University of Alberta

Operation of a Membrane Bioreactor In a Biological Nutrient Removal Configuration

by

Geoffry Rodney Heise ©

A thesis submitted to the Faculty of Graduate Studies and Research in partial fulfillment of the requirements for the degree of Master of Science

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For my family,
Melody and Alexander Joel

Abstract

Hollow fibre membrane bioreactor (MBR) technology offers municipalities an attractive alternative to conventional activated sludge systems with gravity clarification. Three bench-scale 80 litre MBRs were operated in a combined biological nutrient removal (BNR) configuration over a 75-day period. The MBRs were operated in parallel at a fixed mixed liquor suspended solids (MLSS) concentration of 10,000 mg/L at varying hydraulic retention times (HRT) of 12, 9 and 6 hours. Less than detection limit levels of total suspended solids (TSS) and 5-day biochemical oxygen demand (BOD_5) were consistently observed in the permeate over the 75-day study period for all three reactors, regardless of loading. Total nitrogen removal was greater than 60% with complete removal (99.6%) of ammonia. Due to unstable and frequently low organic loading rates, very high sludge residence times (SRT) and poor denitrification, total phosphorus removal was lower than expected for a full BNR system. Average permeate total phosphorus concentrations ranged from 2 to 3 mg/L, although some permeate total phosphorus results of less than 1 mg/L were observed at a 6 hour retention time. With low permeate flux rates less than 15 L/m^2h and average suction pressure requirements less than 5 kPa, very little fouling of the 0.04 μm pore size hollow fibre membranes was observed.

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From the time I commenced my Master's program in 1999 to its completion in 2002, many splendid events have occurred in my life. I reached the milestone of 30 years of age. My dog and my shadow, Murphy, joined our family. My precious son, Alexander Joel was born. Balancing family life, a career, and academics has indeed been a struggle, but I made it. All of this would not have been possible if were not for the enduring support of my wife, Melody, my parents and siblings, and the Larsen family. Melody has been very patient with me the last few years; the time has come for me to focus on her and my little A.J. and leave this thesis behind for others to learn from and build upon.

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Table of Contents

1 Introduction 1
2 Background and Theory 3
  2.1 Gold Bar Wastewater Treatment Plant 3
  2.2 Total Load Limit 4
  2.3 Wastewater Reuse 5
  2.4 The BNR Process 5
    2.4.1 Historical Nutrient Removal 6
    2.4.2 Biological Phosphorus Removal 6
    2.4.3 Nitrogen Removal 8
      2.4.3.1 Nitrification 8
      2.4.3.2 Denitrification 10
    2.4.4 Sludge Residence Time and the BNR Process 11
      2.4.4.1 SRT and Temperature Effects 13
    2.4.5 Bioreactor Configuration 14
    2.4.6 Clarification-The Limiting Step 15
  2.5 Membranes 16
    2.5.1 Definition of Terms 16
    2.5.2 Element or Module Configurations 18
  2.6 Applications of Membrane Processes In Municipal Wastewater Treatment 19
  2.7 Membrane Bioreactors 19
    2.7.1 Recirculated MBRs 22
    2.7.2 Integrated or Submerged MBRs 26
3 Experimental Design 32
  3.1 Experimental Set-up 32
  3.2 Effect of HRT on the BNR process 32
  3.3 Hollow Fibre Membranes 33
  3.4 Equipment Design and Setup 37
    3.4.1 MBR-Feed System 37
    3.4.2 Reactor Size 39
    3.4.3 Liquid Level Control 43
      3.4.3.1 Foam Control 43
    3.4.4 Membrane Pump 44
    3.4.5 Mixed Liquor Recycle 45
    3.4.6 Aeration 46
    3.4.7 Mixing 49
    3.4.8 Online Monitoring 50
      3.4.8.1 Dissolved Oxygen Monitoring 50
      3.4.8.2 Temperature Monitoring 50
      3.4.8.3 Pressure Monitoring 51
  3.5 Sampling Design 51
  3.6 Analytical Methods 52
    3.6.1 TSS determination 52
    3.6.2 VSS determination 53
    3.6.3 MLSS/MLVSS determination 53
    3.6.4 BOD_5 determination 54

3.6.5 Ammonia Determination 56
3.6.6 Total Oxidized Nitrogen Determination 56
3.6.7 Total Kjeldahl Nitrogen Determination 56
3.6.8 Total Phosphorus Determination 56
3.6.9 COD Determination 56
3.6.10 Dissolved Organic Carbon 57
3.6.11 Turbidity 57
3.6.12 pH 57
3.6.13 Coliforms 57
3.6.14 Volatile Fatty Acid Determination 57
3.6.15 Reactor Profile Analyses 57
4 Results and Discussion 59
4.1 Reactor Control 59
4.1.1 Solids Maintenance 59
4.1.1.1 Mixed Liquor Solids Concentration 59
4.1.1.2 SRT 63
4.1.1.3 pH 64
4.1.1.4 Organic Loading 65
4.1.2 Membrane Performance 66
4.1.3 Membrane Fouling 69
4.1.3.1 Fouling Control 69
4.1.3.2 Chemical Cleaning 71
4.1.3.3 Temperature Effects 72
4.2 Permeate Quality 72
4.2.1 Solids Removal 72
4.2.2 Microorganism Reduction 75
4.2.3 Carbon Reduction 77
4.2.3.1 5 Day Biochemical Oxygen Demand 77
4.2.3.2 Chemical Oxygen Demand 77
4.2.3.3 Dissolved Organic Carbon 80
4.2.4 Nitrogen Reduction 82
4.2.5 Phosphorus removal 88
4.2.6 MBR Profiling 91
5 Conclusions 96
6 Recommendations 98
References 100
Appendix A - MBR Control Software 105
Appendix B - Pressure Transducer Calibration 112

List of Figures

Figure 1. The effect of temperature on the minimum aerobic SRT for nitrogen and phosphorus removal (after Grady *et al.*, 1999). 14

Figure 2. The three-stage Bardenpho process with Johannesburg modification process utilized at GBWWTP. 15

Figure 3. Recirculated MBR. 21

Figure 4. Integrated or submersed MBR. 21

Figure 5. Hollow fibre membranes immersed in external tank after Husain and Coté (1999). 29

Figure 6. Hollow fibre membranes immersed in bioreactor after Husain and Coté (1999). 29

Figure 7. Outside-in flow pattern of hollow fibre membrane. 35

Figure 8. (a) ZW-10 element supplied by Zenon Environmental Inc. Membrane surface area is equal to 0.929 m^2. (b) Cross-section of membrane element aeration tube and resulting bubble pattern. 36

Figure 9. MBR project configuration. 38

Figure 10. Plan view of reactor 40

Figure 11. Profile view of reactor. Blocked out portion of reactor was not utilized and only acted as a support structure to the four upper chambers. 40

Figure 12. Three dimensional depiction of reactor and corresponding cell volumes. 41

Figure 13. Three-dimensional wireframe reactor illustrating flow pattern. 42

Figure 14. Plan view of reactor showing the flow pattern between individual cells. 42

Figure 15. Plan view of MBR mixed liquor recycle configuration. 46

Figure 16. Fine bubble aerator. 47

Figure 17. Mixing pattern created by fine bubble aerator. 48

Figure 18. (a) Original mixing shaft and impeller. (b) Modified mixing shaft and impellers. 49

Figure 19. Mixed liquor suspended and volatile solids maintenance for MBR 1 62

Figure 20. Mixed liquor suspended and volatile solids maintenance for MBR 2 62

Figure 21. Mixed liquor suspended and volatile solids maintenance for MBR 3 63

Figure 22. Mixed liquor pH results for MBR 1, MBR 2 and MBR 3. 64

Figure 23. Average suction pressure requirements for MBR 1, MBR 2 and MBR 3. 67

Figure 24. Snapshot of 80 minute permeate production period for MBR 3 on day 59. 71
Figure 25. MBR-Feed TSS results. 74
Figure 26. Permeate TSS concentration for MBR 1, MBR 2 and MBR 3. 74
Figure 27. Permeate turbidity results for MBR 1, MBR 2 and MBR 3. 75
Figure 28. Five-day BOD results for the MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate. 78
Figure 29. MBR-Feed COD concentration. 79
Figure 30. Permeate COD concentration for MBR 1, MBR 2 and MBR 3. 79
Figure 31. Dissolved organic carbon results for MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate. 81
Figure 32. MBR-Feed TKN concentration. 84
Figure 33. Permeate TKN concentration for MBR 1, MBR 2 and MBR 3. 84
Figure 34. MBR-Feed ammonia concentration. 85
Figure 35. Permeate ammonia concentration for MBR 1, MBR 2 and MBR 3. 85
Figure 36. Total oxidized nitrogen (NO_2^-+NO_3^- = TOXN) concentration for MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate. 86
Figure 37. Total phosphorus concentration distribution for MBR 1, MBR 2 and MBR 3 permeate. 89
Figure 38. MBR 1, MBR 2, and MBR 3 total phosphorus removal as compared to MBR-Feed soluble COD concentration. 90
Figure 39. Mixed liquor suspended and volatile solids MBR profile. 93
Figure 40. Ammonia MBR profile. 94
Figure 41. Nitrate MBR profile. 94
Figure 42. Ortho-phosphate MBR profile. 95
Figure 43. Membrane control software "Main Start" screen. 109
Figure 44. "Select Job" screen. 109
Figure 45. "New Job Name" screen. 110
Figure 46. "Main Control" screen. MBRs 1, 2 and 3 operated in manual mode. Feed valve and permeate pump are "Off". 110
Figure 47. "Main Control" screen with MBR1, 2 and 3 operated in "Automatic" mode. 111
Figure 48. Calibration results for pressure transducer #1. 113
Figure 49. Plot of pressure (kPa) versus volts for pressure transducer #1. Linear equation was generated using Microsoft® EXCEL's linear regression tool. 113

Figure 50. Calibration results for pressure transducer #2. 114

Figure 51. Plot of pressure (kPa) versus volts for pressure transducer #2. Linear equation was generated using Microsoft® EXCEL's linear regression tool.......... 114

Figure 52. Calibration results for pressure transducer #3. 115

Figure 53. Plot of pressure (kPa) versus volts for pressure transducer #3. Linear equation was generated using Microsoft® EXCEL's linear regression tool.......... 115

List of Tables

Table 1. GBWWTP average licensed parameter concentrations for 2001 in comparison to current limits for licensed parameters and the new limits for 2005 as stipulated by Alberta Environment License to Operate Approval No. 95-MUN-117. 4

Table 2. Comparison of membrane module configurations after Paulson (1995). 34

Table 3. Theoretical flux needed to achieve desired HRT 39

Table 4. Relationship of float position to MBR primary effluent feed system 43

Table 5. Sampling schedule. MBR-Feed "Routine" analyses include: TSS, BOD_5, COD, soluble COD, DOC, VFA, TKN, NH_3N, TOXN, and TP. MBR1, 2, 3 "Routine" analyses include: TSS, BOD_5, $CBOD_5$, COD, DOC, TKN, NH_3N, TOXN, TP, and Turbidity. "Profile" analyses include: HACH NH_3N, NO_3^-N, and ortho-phosphate. MLSS sample was analyzed for TSS, VSS and pH. 55

Table 6. HACH analytical methods used for zone profiling. 58

Table 7. Summary of operating conditions for MBR 1, MBR 2 and MBR 3. 61

Table 8. Typical operating conditions for activated sludge systems. Taken from Metcalf and Eddy Inc. (1991). 65

Table 9. Clean water membrane permeability results. 66

Table 10. Summary of actions taken that may have affected suction pressure requirements for MBR 1, MBR 2 and MBR 3. 68

Table 11. Permeate quality results for MBR 1, MBR 2, and MBR 3 as compared to the MBR-Feed. 73

Table 12. Total and fecal coliform results (CFU/100 mL) for MBR-Feed, MBR 1, MBR 2 and MBR 3. 76

Table 13. Example of online data collected for MBR. 111

1 Introduction

Regulatory bodies are tightening the effluent discharge criteria for pollution loads entering North America's surface waters. This has driven municipalities to investigate new advanced wastewater treatment technologies to meet these pollution reduction demands. Membrane technology offers a physical separation process that is compact, easy to operate and maintain, and provides consistent high quality effluent regardless of the feed characteristics. The membrane bioreactor (MBR), in combination with biological nutrient removal (BNR) processes, is becoming a practical replacement to conventional activated sludge (CAS) designs.

MBRs, utilizing hollow fibre membrane elements, offer an attractive alternative to the CAS/secondary clarification system. Because the membrane acts as a physical barrier to solids, some larger molecular weight organics, and microorganisms, secondary clarification is not required to separate mixed liquor suspended solids (MLSS) from the treated wastewater. Limitations with the operation of a secondary clarifier, such as sludge bulking or rising sludge due to denitrification, are no longer experienced with MBR systems. The hollow fibre membrane provides consistent removal of constituents from wastewater regardless of influent characteristics or the settling properties of the activated sludge. An MBR is typically operated at an MLSS concentration of between 10 and 30 g/L. At this MLSS concentration, the removal of ammonia and total nitrogen is enhanced. The high MLSS also has the direct impact of reducing the hydraulic retention time (HRT) for existing wastewater treatment facilities and reducing the overall footprint of new MBR facilities. Chemical phosphorus removal has been proven successful for both pilot and full-scale MBR systems. Little research has investigated chemical-free biological mediated phosphorus removal using MBR technology. It is within the scope of this study to investigate combined BNR (both nitrogen and phosphorus removal) using a bench-scale MBR system.

Three bench-scale MBRs were designed, constructed, and operated over a four-month period. Configured for combined biological phosphorus (Bio-P) and nitrogen removal, the MBRs were operated under different loading conditions to evaluate the effluent quality and process stability of this novel technology.

The objectives of this project were twofold:

1. to design and operate a bench-scale continuous flow BNR unit incorporating a MBR system; and
2. to evaluate the performance of the BNR-MBR system in terms of TSS, BOD_5, nitrogen and phosphorus reduction.

2 Background and Theory

2.1 Gold Bar Wastewater Treatment Plant

The Gold Bar Wastewater Treatment Plant (GBWWTP) is located on the south bank of the North Saskatchewan River in Edmonton, Alberta. The facility has a primary treatment capacity of 910 million litres per day (ML/day), and treats on average 310 ML/day of wastewater, with a peak secondary treatment capacity of 420 ML/day. With the retrofit completed in early 2002, the GBWWTP is now a fully combined BNR facility. The upgrade was facilitated by Alberta Environment's (AE) new approval to operate for GBWTTP. Effective June 2005, the new approval will require GBWTTP to meet total phosphorus concentration levels of less than 1 mg/L and ammonia limits of less than 10 mg/L in the winter (November to April) and less than 5 mg/L in the summer (May to October). With biological phosphorus and nitrogen removal and a state-of-the-art ultraviolet (UV) microorganism reduction system capable of removing 5 log fecal coliform bacteria, GBWWTP produces excellent quality effluent on a consistent basis and should easily meet the new approval limits. Table 1 compares the existing limits, new 2005 limits, and 2001 effluent quality.

Table 1. GBWWTP average licensed parameter concentrations for 2001 in comparison to current limits for licensed parameters and the new limits for 2005 as stipulated by Alberta Environment License to Operate Approval No. 95-MUN-117.

Parameter	Current Limit	2005 Limit	2001-Average Concentration
5-day Carbonaceous BOD	≤ 25.0 mg/L	≤ 20.0 mg/L	5.7 mg/L
Total Suspended Solids	≤ 25.0 mg/L	≤ 20.0 mg/L	9.5 mg/L
Fecal Coliforms	≤ 200 colony forming units (CFU) per 100 mL	≤ 200 CFU/100 mL	31 CFU/100 mL
Total Phosphorus	No Limit	≤ 1 mg/L	1.0 mg/L
Ammonia-N (Winter)	No Limit	≤ 10.0 mg/L	9.5 mg/L
Ammonia-N (Summer)	No Limit	≤ 5.0 mg/L	6.4 mg/L

2.2 Total Load Limit

Recently, AE has discussed the implementation of a total maximum load limit (TML) for all point sources entering the North Saskatchewan River from the Edmonton region. This includes storm and combined sewer outfalls, plant and secondary bypasses, and effluent from GBWWTP. Therefore, not only must GBWWTP meet the concentration limit for a pollutant, but may also have to meet a licensing requirement for the total yearly load of a pollutant entering the river from all point sources. Even with the retrofit to combined BNR, GBWWTP may not be able to reduce total loading to the river. Alternative treatment technology must be investigated to improve pollution reduction for all licensed parameters. MBR technology may offer an attractive alternative to plant expansion to meet these future demands.

2.3 Wastewater Reuse

In addition to the pollution reduction demands placed on Alberta's existing and future wastewater treatment facilities, the opportunity has arisen to investigate wastewater effluent reuse. Without excellent treatment, North American fresh water resources, including surface waters and ground water aquifers, could rapidly become polluted and depleted due to industrial and population growth. With the enforcement of stringent guidelines from regulatory bodies and advances in wastewater treatment technology, effluent quality has steadily improved to a level where wastewater effluent reuse is a feasible and practical concept.

Industry surrounding GBWWTP has shown a great interest in the effluent the wastewater treatment facility produces. If biological activity can be eliminated ~~and~~ pollutants such as suspended solids and ammonia can be reduced to non-detectable levels, the effluent could be well suited for process water applications. The municipality has also voiced interest in using the effluent for landscape irrigation of golf courses, parks, and freeways and as a water supplement to local ponds. Wastewater reuse may be a less expensive alternative to the use of water from the potable water distribution system. Future investigation into wastewater reuse is essential to establish its feasibility and to determine health and environmental impacts.

2.4 The BNR Process

The discharge of effluents high in phosphorus and nitrogen compounds from wastewater treatment plants has become a tremendous environmental and legal concern. Phosphorus, a major component in commercial soaps and detergents, is a major environmental concern as it causes algae to grow in receiving waters via a process called cultural or anthropological eutrophication. Algae, when it dies and settles, reduces the dissolved oxygen (DO) concentrations of lakes and rivers and can cause the extinction of higher organisms within ecosystems. The oxidation of nitrogen compounds can also reduce the available oxygen for aquatic life. Human wastes and industrial wastes from slaughterhouses or food processing plants have a direct impact on the concentration of nitrogen compounds in wastewater. These effluents discharged into the sewer system are very high in protein. Protein is easily metabolized or chemically broken down to release ammonia, carbon dioxide, and water during wastewater treatment. Ammonia is a strong base and dissolves lipids in tissues making it extremely toxic to fish and other higher aquatic organisms. Other nitrogen compounds found in wastewater influent, such as nitrates and nitrites, pose a human health concern; therefore, they limit the reuse potential of wastewater effluents unless they are removed. Strict government licensing requirements have forced treatment plants to upgrade to alternative and cost-effective wastewater treatment systems in order to reduce these pollutants.

BNR can be described in general terms as the natural, biologically mediated reduction of nitrogen and/or phosphorus by the microorganisms found in activated sludge. By manipulating the environmental conditions of the activated sludge, one can optimize the growth conditions of specialized bacteria responsible for nitrogen or phosphorus reduction.

2.4.1 Historical Nutrient Removal

In the past, physical and chemical processes for the reduction of nitrogen and phosphorus were used extensively in wastewater treatment. These processes are still used today as stand-alone processes or as back-up systems to BNR systems. The chemical precipitation of phosphorus with aluminum sulfate has been an effective process for the reduction of phosphorus; however, the disposal of aluminum-contaminated sludge has become a public health concern. Ferric chloride is another popular chemical that can be used to precipitate phosphorus from activated sludge or secondary effluent. Ultimately, the high costs of chemicals limit their use in wastewater treatment applications. Alternative nutrient removal methods such as ammonia stripping and ion exchange processes have extremely high capital and operating costs and are not practical for large wastewater treatment systems (Metcalf and Eddy Inc., 1991).

2.4.2 Biological Phosphorus Removal

Biological phosphorus (Bio-P) removal can be considered to be a two-stage metabolic process with both the Comeau-Wentzel model and the Mino models being accepted as the biochemical models for this process (Arun *et al.*, 1988; Comeau *et al.*, 1986; Mino *et al.*, 1998; Wentzel *et al.*, 1986). Under anaerobic conditions (in the absence of molecularly bound oxygen) heterotrophic phosphorus removing bacteria (Bio-P bacteria) release phosphorus into the mixed liquor as a result of the uptake and storage of short-chain volatile fatty acids (VFA). The most common VFA is acetic acid, and is the easiest metabolized. Others include propionic and butyric acid. These readily available carbon sources are stored intracellularly as poly-β-hydroxybutyrate (PHB). When exposed to aerobic conditions, the Bio-P bacteria utilize the stored PHB as a carbon source for growth. In doing so, the bacteria uptake phosphorus from the mixed liquor and store it as long polyphosphate molecules. In its conversion from a soluble form (ortho-phosphate) to a polymeric molecule stored within the cell biomass (polyphosphate), phosphorus is ultimately removed from wastewater by the wasting of biomass from the bioreactor.

It has been historically believed that the genus *Acinetobacter* was the primary organism responsible for Bio-P removal (Fush and Chen, 1975). Wagner *et al.* (1994) later found that this genus only plays a minor role in the removal of phosphorus from wastewater. Other organisms that were found in large populations in Bio-P removing activated sludge were *Pseudomonas* and *Aeromonas* bacteria. Some of these bacteria were found to also have the capability of denitrification.

Some of the factors affecting Bio-P removal include feed characteristics, temperature and operational conditions. Factors that influence phosphorus release in the anaerobic zone include the presence of nitrates, pH, excess VFA, temperature and easily biodegradable carbon in the influent. Nitrates in the mixed liquor recycle have detrimental effects on the release of phosphorus in the anaerobic zone. Denitrifiers present in the anaerobic zone out-compete Bio-P bacteria for VFA, resulting in denitrification (Kuba *et al.*, 1994). An increase in the ratio of biochemical oxygen demand to total phosphorus in the influent enhanced phosphorus removal (Morales *et al.*, 1991; Qasim *et al.*, 1996). An increase in pH was found to be a contributing factor to the increased release of phosphorus in the anaerobic zone (Pramanik *et al.*, 1997). Recently, some research in the laboratory has shown that phosphate removal has been increased by up to 143% if the operational pH is held between the range of 5.5 to 6.5 for some isolated cultures from activate sludge samples. This is contrary to the belief that the optimal pH range for Bio-P removal is between 7.2 and 7.7 (McGrath *et al.*, 2001). Increasing temperature in the anaerobic zone seemed to enhance phosphorus release in this zone. Mamais and Jenkins (1992) found an optimal temperature range of 28 to 33°C whereas Brdjanovic *et al.* (1997) determined that phosphorus release continually changed between 5 and 30°C with a maximum rate at 20°C. Operational factors of Bio-P removal will be further discussed in section 2.4.4.

2.4.3 Nitrogen Removal

The removal of nitrogen occurs through the processes of nitrification and denitrification. The process of nitrification occurs under aerobic conditions and requires free oxygen molecules. Autotrophic bacteria convert ammonia to nitrite and then nitrate utilizing oxygen as a terminal electron acceptor and consuming alkalinity. Denitrification is the process by which nitrate is converted to molecular nitrogen with the production of alkalinity. Under anoxic conditions (in the absence of dissolved molecular oxygen), heterotrophic bacteria utilize nitrate as a terminal electron acceptor for the metabolism of readily biodegradable organic carbon. Due to its poor solubility in water, nitrogen gas produced under anoxic conditions is released into the atmosphere.

2.4.3.1 Nitrification

Nitrification is the biochemical process by which the ammonium ion (NH_4^+) is oxidized to nitrate (NO_3^-). Under typical wastewater conditions ($6.5 \leq pH \leq 8.0$) aqueous ammonia (NH_3) exists primarily as ammonium ions. Since molecular oxygen, O_2, is used as the terminal electron acceptor, nitrification is an aerobic process. The oxidation of ammonium to nitrate is a two-step process: the first step is the oxidation of ammonium to nitrite (termed nitrosification); the second step is the oxidation of nitrite to nitrate (termed nitrification). Two individual genera of bacteria are responsible for the overall nitrification process. These are *Nitrosomonas* and *Nitrobacter*. Both genera of bacteria are strict aerobes and will not grow in the absence of oxygen. *Nitrosomonas* and

Nitrobacter are considered lithotrophic autotrophs. Lithotrophic organisms are capable of utilizing inorganic sources of energy as electron donors (ammonium (NH_4^+) or nitrite (NO_2^-) are two examples). Autotrophs use only carbon dioxide (CO_2) as a carbon source while heterotrophs need organic carbon, such as acetate (CH_3COO^-) or glucose ($C_6H_{12}O_6$) for energy production. The following equations show the two-step oxidative process and the overall reaction for the conversion of ammonium to nitrate (Metcalf and Eddy Inc., 1991):

Step #1:

$$\textit{Nitrosomonas}: 2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 2H_2O + 4H^+$$ **Equation 1**

Step #2:

$$\textit{Nitrobacter}: 2NO_2^- + O_2 \rightarrow 2NO_3^-$$ **Equation 2**

Overall:

$$2NH_4^+ + 4O_2 \rightarrow 2NO_3^- + 4H^+ + 2H_2O$$ **Equation 3**

For every milligram of NH_4^+ oxidized to nitrate, 4.57 mg of O_2 are consumed by the nitrifying bacteria and 7.14 mg of alkalinity (as $CaCO_3$) are destroyed to buffer the reaction. Therefore, in wastewater treatment, there must not only be sufficient O_2 present for nitrification to occur, but a sufficient amount of alkalinity is also required to buffer any pH shifts towards acidic conditions (which would otherwise be lethal to the nitrifying bacteria). It was reported that the optimal pH range for nitrification occurs between 7.0 and 8.5 (Rabinowitz *et al.*, 1990). In addition to low pH inhibiting nitrification, low temperature has a profound limiting effect on the growth of nitrifying bacteria. Heavy metals and some organic molecules have also been found extremely toxic to *Nitrosomonas* and *Nitrobacter,* as well as most other bacteria present in activated sludge systems.

2.4.3.2 Denitrification

Denitrification is the process by which oxidized nitrogen (nitrate/nitrite) is converted to intermediate nitrogen gases (NO and N_2O), and ultimately molecular nitrogen, N_2. This process occurs under anaerobic conditions because molecular oxygen is considered inhibitory to the denitrification process. In the presence of oxygen, the necessary enzymes (nitrate reductase and nitrite reductase) responsible for the reduction of NO_3^- and NO_2^- to molecular nitrogen are not synthesized.

For denitrification to occur nitrate or nitrite must be utilized as the terminal electron acceptor in the electron transport chain. This condition is termed anoxic, because oxygen is present only in the form of oxidized nitrogen (NO_3^-/NO_2^-). Van Haandel *et al.* (1981) stated that there are four conditions necessary for denitrification to occur. First, nitrate must be present. This implies nitrate is the prerequisite for denitrification. Second, dissolved oxygen must be absent. This satisfies the anoxic conditions needed for denitrification. Third, a bacterial population must be present that can utilize nitrate and oxygen as terminal electron acceptors; therefore, a bacterial mass must be able to adapt biochemically to both aerobic and anoxic conditions (facultative anaerobes). *Escherichia coli* and *Pseudomonas sp.* are both facultative heterotrophic bacteria capable of partial and complete denitrification. Both organisms are found in high populations in activated sludge along with many other genera of denitrifying bacteria: *Achromobacter, Aerobacter, Alcaligenes, Bacillus, Flavobacterium, Lactobacillus, Micrococcus, Proteus, and Spirillum* (Metcalf and Eddy Inc., 1991). Finally, a suitable electron donor must be present. The electron donor or "energy source" can be present in any one of three forms. First, the energy source can be an external source found in wastewater influent, such as methanol (CH_3OH). Second, the energy source can be produced internally in the wastewater stream from other chemical and biochemical processes, such as VFA produced from anaerobic fermentation. Finally, the energy source can be produced from the death and lysis of other microorganisms in the wastewater stream. The following reaction shows the overall conversion of nitrate to molecular nitrogen:

Step #1

$$6NO_3^- + 2CH_3OH \rightarrow 6NO_2^- + 2CO_2 + 4H_2O$$ **Equation 4**

Step #2

$$6NO_2^- + 3CH_3OH \rightarrow 3N_2 + 3H_2O + 3CO_2 + 6OH^-$$ **Equation 5**

Overall:

$$6NO_3^- + 5CH_3OH \rightarrow 5CO_2 + 3N_2 + 7H_2O + 6OH^-$$ **Equation 6**

Van Haandel *et al.* (1981) showed that for every one milligram of nitrate converted to nitrogen, 2.86 mg of oxygen is not utilized for the removal of organic matter by activated sludge. This represents a 63% recovery of the aeration needed for nitrification. The oxygen saved may be as much as 15 to 20% of the total oxygen requirements of an activated sludge treatment system. Rabinowitz *et al.* (1990) has suggested that aeration savings may be even higher due to the inefficiencies associated with conventional aeration equipment. Alkalinity production is also a result of nitrification with a net recovery of 3.57 mg of alkalinity for each milligram of nitrate denitrified. This amounts to a 50 % recovery of alkalinity that was lost in the nitrification step.

2.4.4 Sludge Residence Time and the BNR Process

Sludge residence time (SRT) represents the average length of time a particulate constituent stays in the bioreactor. For a CAS system where all solids are retained in the bioreactor, SRT is defined by equation 7. When the reactor volume is set, the wasting flow rate, Q_W, governs the SRT of the bioreactor. SRT is used as the main operating parameter of a CAS system. By controlling the SRT, one can control the mass of microorganisms in the reactor and ultimately the amount of substrate removal. The limiting restraint on SRT in a CAS is that at an increased SRT, the settling characteristics of the activated sludge degrade rapidly resulting in poor effluent quality from a secondary clarifier. Since sludge-settling characteristics are not important in an MBR system, the SRT can be operated at greater than 50 days with no difficulties and minimal sludge production. The following equation shows the relationship between SRT, reactor volume, and waste stream flow:

$$\theta_c = \frac{V}{Q_w}$$

Equation 7

where: θ_c = SRT or mean cell residence time, T

V = reactor volume, L^3

Q_w = waste stream flow, L^3T^{-1}

The effluent substrate concentration of a bioreactor is governed by the SRT of the biomass in the reactor. If the bioreactor volume is fixed and the flow is increased at a constant SRT, the mass of microorganisms in the bioreactor must increase to keep consistent with the mass of substrate being removed. As a consequence, the biomass concentration must increase proportionally. If the flow or organic loading is changed for any reason, the concentration of biomass will also change to maintain the mass of organisms sufficient to produce an effluent substrate concentration consistent with the SRT. If the influent substrate concentration is changed, the active biomass will change until the effluent substrate concentration is consistent with the SRT. This will make the substrate concentration of the effluent independent of the influent substrate concentration.

SRT is one of many important factors in the operation of a combined BNR system. The SRT of the individual zones of the bioreactor (anaerobic, anoxic, and aerobic) plays a more important role in a combined BNR system than the total SRT. Organism selection is affected by the aerobic SRT. At a longer SRT, specialized bacteria such as nitrifiers and Bio-P bacteria can be sustained in a system. Mamais and Jenkins

(1992) showed that Bio-P removal functioned efficiently at an aerobic SRT greater than 2.9 days for temperatures between 13.5 and 20°C. Brdjanovic *et al.* (1998b) achieved good Bio-P removal at a total SRT (anaerobic and aerobic) of 8, 16, and 32 days at temperatures of 20, 10, and 5°C, respectively. Minimum aerobic SRT for nitrification have been found to be much higher than Bio-P removal due to the slow growth of nitrifying organisms. The SRT to achieve nitrification will therefore be the controlling factor of a combined BNR process. The minimum aerobic SRT at which washout will occur is typically 1 day (at 20°C) to 4 days (at 8°C) longer than the SRT required for Bio-P bacteria. The anaerobic SRT should be at least 0.5 days (Grady *et al.*, 1999). Typically, Bio-P removal processes require an SRT of 1 to 1.5 days depending on temperature. An anaerobic SRT in the range of 2 to 4 days produces better phosphorus removal due to the fermentation of soluble substrate into volatile fatty acids under extended anaerobic conditions (Matsuo, 1994). At GBWWTP, fermentation in the anaerobic zone is not very important due to the recent commissioning of a primary sludge fermenter which provides a constant source of VFA for the both the pre-anoxic denitrification zone and anaerobic phosphorus release zone. A constant flow of VFA rich industrial wastewater also goes to the wastewater treatment facility from a local chemical plant. Anoxic zone SRT does not have to be very long provided the zone is supplied with sufficient nitrate for denitrification. Most combined BNR systems are designed for an anaerobic plus anoxic SRT of at least 2 to 3 days. If hydrolysis of slowly biodegradable organic matter is desired, then a higher SRT is required (Grady *et al.*, 1999).

2.4.4.1 SRT and Temperature Effects

Temperature is an important factor with respect to the BNR process and overall SRT selection. Temperature affects the kinetics of the biochemical conversions utilized by the bacteria responsible for both phosphorus removal and nitrogen removal. Nitrification is the process most affected by temperature. Temperatures less than 15°C have a strong negative effect on the growth of nitrifying organisms. As illustrated in Figure 1, lower temperatures require a longer SRT to achieve nitrification due to the slow growth of nitrifying organisms. Denitrification is also affected by temperature, and usually requires a longer SRT at lower temperatures to facilitate BOD and nitrate removal. Because low temperatures also affect biological phosphorus release and uptake, a longer SRT is also required in the anaerobic zone to facilitate good phosphorus removal. In addition to temperature, other environmental affects on the BNR process include pH and dissolved oxygen concentrations (Grady *et al.*, 1999).

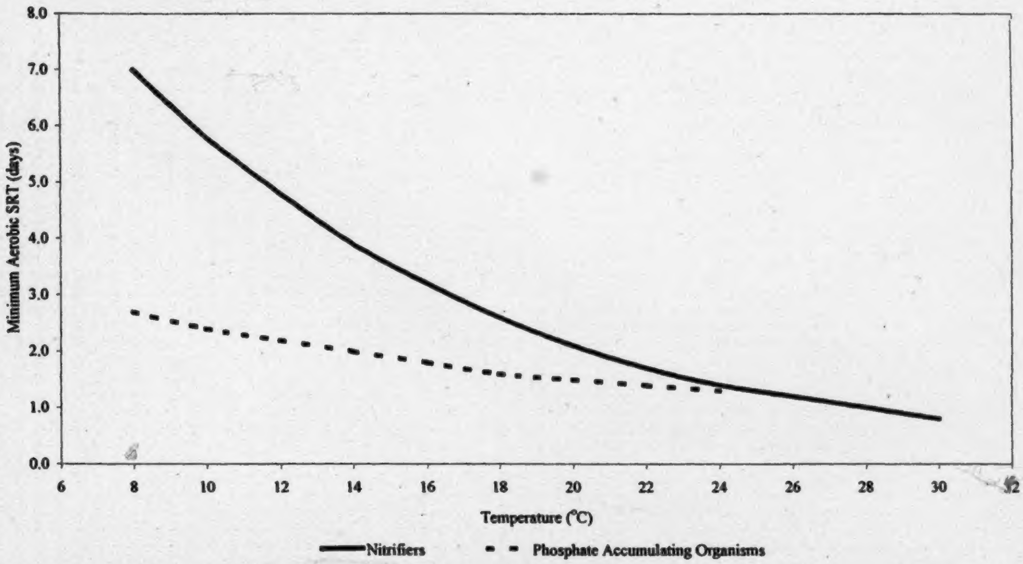

Figure 1. The effect of temperature on the minimum aerobic SRT for nitrogen and phosphorus removal (after Grady *et al.*, 1999).

2.4.5 Bioreactor Configuration

By partitioning sections of the bioreactor into anaerobic, anoxic, and aerobic zones and rerouting or adding sludge recycle lines, most, if not all, plug-flow CAS systems can be easily converted to combined BNR systems. Many configurations of combined BNR systems, such as the A^2/O™, UCT, VIP and four-stage Bardenpho process, have been successfully operated at full-scale.

GBWWTP uses a variation of the A^2/O™ process with the addition of a pre-anoxic zone upstream of the anaerobic zone. Named the three-stage Bardenpho process with the Johannesburg modification, the pre-anoxic zone removes nitrate contained in the return activated sludge (RAS) from the clarifier (Figure 2). Nitrate is detrimental to the biological phosphorus removal process by providing a terminal electron acceptor to heterotrophic bacteria found within this zone, thereby limiting phosphorus release. By directing the RAS to a pre-anoxic zone, nitrate is removed via denitrification. This ensures the conditions necessary for phosphorus release are maintained in the anaerobic zone. The typical recycle rate for the nitrified mixed liquor return is 1 to 3Q. Thickened RAS from the clarifier is typically returned at a rate of 0.75Q to the front of the reactor.

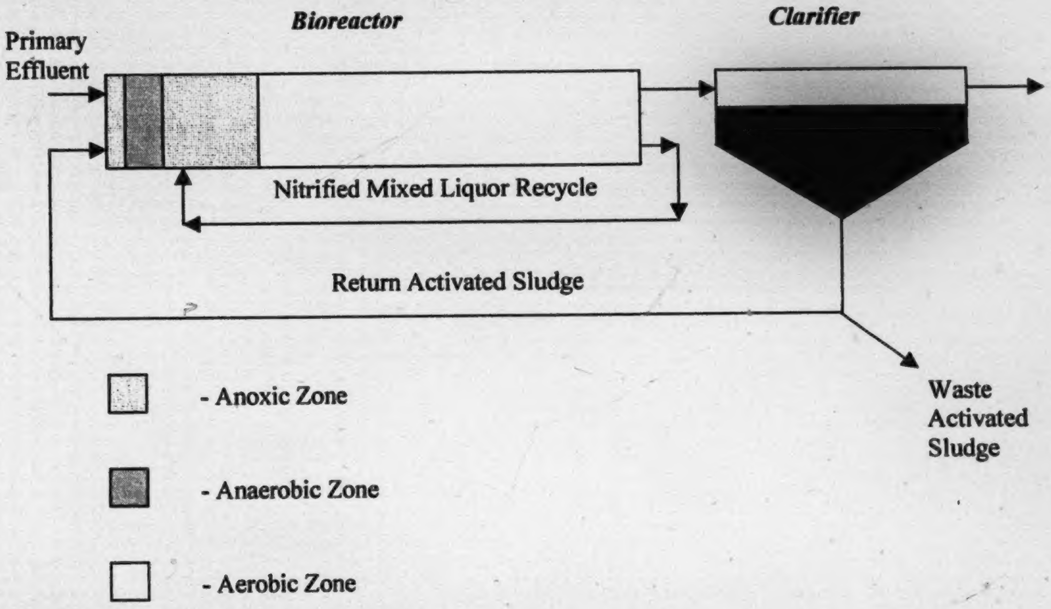
Bioreactor
Clarifier
Primary Effluent
Nitrified Mixed Liquor Recycle
Return Activated Sludge
Waste Activated Sludge
- Anoxic Zone
- Anaerobic Zone
- Aerobic Zone

Figure 2. The three-stage Bardenpho process with Johannesburg modification process utilized at GBWWTP.

2.4.6 Clarification-The Limiting Step

Secondary clarification is the limiting step in the operation and design of a CAS process. Optimal conditions in the bioreactor, such as dissolved oxygen concentration and food-to-mass ratio (F/M ratio) must be maintained in order to promote the growth of flocculating bacteria in the bioreactor and good settling sludge. When optimal conditions are not maintained, the overgrowth of filamentous bacteria will result in poor settling sludge and, ultimately, in high effluent concentrations of licensed parameters.

The secondary clarification step also limits the MLSS that can be maintained in the bioreactor. An activated sludge with a high MLSS will not settle well and will require a longer detention time in the clarifier, either limiting flow entering the clarifier or increasing the required volume of the clarifier. With the variations in organic loading and influent characteristics to a CAS system, maintaining optimal operating conditions is a challenge with the potential for bulking sludge constantly lingering.

2.5 Membranes

2.5.1 Definition of Terms

Membranes are physical, semi-permeable barriers used to separate constituents from a fluid medium. These constituents can be in gas, liquid or solid states. The separation of materials within a medium is a function of their physical and chemical properties as well as membrane pore size and materials. The driving force for mass transfer may be a concentration gradient, electrical potential, or pressure differential (Mallevaille *et al.*, 1996). Membranes are typically classified according to their separation mechanism, geometry, morphology, and materials.

Membrane treatment generally produces two streams: a permeate and a retentate. Permeate is that portion of the feed that passes through the membrane (the treated stream). Retentate is that portion of the feed that does not pass through the membrane and contains the majority of constituents that are removed from the feed (the concentrated stream).

Two types of filtration are commonly practiced: dead-end filtration and crossflow filtration. In dead-end filtration, the feed flow is perpendicular to the membrane or filter surface; this results in the build-up of a cake layer and a subsequent decline in specific flux, with no generation of a concentrate or retentate stream. This type of filtration resembles that of the classic filtration practices used in water treatment, such as rapid sand filtration. Membranes used in this configuration are either discarded after severe flux decline or cleaned by backwashing (permeate or air flow reversal). Crossflow filtration limits the formation of a cake layer by passing the feed flow parallel (or tangential) to the membrane surface. At the applied operating pressures, turbulent flow is achieved which scours the surface of the membrane and limits the decline in flux by continually removing surface deposits. Crossflow filtration greatly extends the productivity of the membrane in comparison to dead-end filtration; however, energy requirements are much greater for crossflow filtration (Rautenbach and Albrecht, 1989).

Membrane performance is often measured by permeate quality as well as permeate flux. The term "flux" is used to describe the movement of moles, mass, or volume of a specified constituent through a unit of membrane surface area per unit time. Units commonly used in the current literature to express flux include [m^3/m^2h] or [L/m^2h] (Koros *et al.*, 1996). Flux is function of transmembrane pressure, viscosity of the fluid medium, and the apparent membrane resistance. For clean water passing through a clean membrane, flux can be described by Equation 8.

$$J = \frac{\delta P}{\mu R_m}$$

Equation 8

where, J = flux (m^3/m^2s)

δP = transmembrane pressure (N/m^2)

μ = absolute viscosity (Ns/m^2)

R_m = apparent membrane resistance (1/m)

Another common measure of membrane performance is the term "percent recovery". Percent (%) recovery is defined as the permeate flow exiting the membrane module divided by the feed flow entering the module.

The cost of replacing membranes, as reflected in the membrane's life span, is often the single largest component of operating cost. A higher permeate flux will correspond to lower operating costs by reducing the amount of membrane area to be replaced. Thus, permeate flux and the factors that influence it are central considerations in determining membrane performance and cost. The major factors that affect permeate flux include pressure difference across the membrane (transmembrane pressure), temperature, feed chemistry, membrane surface charge and membrane fouling. Permeate flux, as well as operational costs, increase proportionately with the increase in transmembrane pressure. As temperature increases, viscosity of permeate water decreases, corresponding to an increase in permeate flux (Wiesner and Aptel, 1996).

Fouling is the largest contributor to the decline of permeate flux for a membrane. The deposition and accumulation of suspended or dissolved constituents on the surface or within the structure of the membrane result in flux decline, which ultimately determines the cleaning frequency of the membrane and its life span. A good membrane material must produce high specific flux (high permeability), reject targeted contaminants (high selectivity), resist fouling, have high mechanical and thermal stability, and have some degree of chemical resistance (DiGiano *et al.*, 1999).

2.5.2 Element or Module Configurations

Elements, commonly referred to as modules, are the engineered operational components housing the membrane material. The element is composed of the membranes, pressure support structures, feed inlet and concentrate outlet ports, and permeate draw-off points. There are many points that must be considered in the design of a membrane element (Aptel and Buckley, 1996). These include:

- ensuring sufficient circulation of the fluid to be treated in order to prevent plugging and accommodate cleaning,
- producing a compact module; that is, to achieve maximum active membrane surface per unit volume,
- avoiding leakage between the feed and the permeate compartments,
- ease of cleaning (hydraulic, chemical, sterilization), and
- ease of assembly and disassembly.

2.6 Applications of Membrane Processes In Municipal Wastewater Treatment

Pressure-driven membrane processes, where the driving force for mass transfer is a pressure differential, can be classified into four general separation processes: microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), and reverse osmosis (RO). MF is used primarily for the removal of suspended particulate matter in the macromolecular range of greater than 0.1 microns (µm) in size. MF produces higher permeate flux than the other membrane processes, with RO typically having the lowest permeate recovery. UF is capable of removing large molecular weight organic molecules and tiny colloidal matter less than 0.1 µm. This level of treatment can provide partial disinfection of water with the removal of bacteria, protozoa, and some viruses. Typically, transmembrane pressures of 170 kiloPascals (kPa) or greater are required for MF and pressures as high as 1,000 kPa have been utilized for UF. NF and RO are diffusive membrane processes capable of producing high quality permeate with the selective removal of most aqueous salts, metal ions, and small molecular weight organic compounds such as sugars. RO is high-pressure separation process (greater than 3,500 kPa) and is capable of producing water that meets drinking water standards under nearly all conditions.

Two applications of pressure-driven membrane processes have been used in advanced wastewater treatment. The first application, tertiary "end of pipe" treatment, has been used extensively as a polishing step for secondary effluent reclamation and reuse. For an extensive literature review on tertiary wastewater treatment applications of membrane technology, refer to the Technical Report by Heise *et al.* (2000). The second application, the MBR, has allowed WWTPs to increase plant removal efficiencies and produce better effluent quality than conventional treatment by eliminating secondary clarification.

2.7 Membrane Bioreactors

Smith *et al.* (1968) were the first researchers to propose the use of membranes to separate secondary effluent from the MLSS. This separation technology has since become known as the MBR. The MBR has two original properties related to the use of membrane separation (Huyard *et al.*, 1992). First, the effective retention of all microorganisms ensures a high degree of disinfection of the treated water, allows the hydraulic detention time to be independent of the sludge age, and results in a high biomass concentration in the bioreactor (Buisson *et al.*, 1998; Cicek, 1998). Second, the retention of higher molecular weight organic particulate and soluble compounds keeps slowly biodegradable molecules within the bioreactor while non-biodegradable constituents are discharged with the sludge rather than with the treated water (Cicek, 1998). Membrane filtration serves as a safety unit in case of bioreactor malfunction, because membrane filtration alone can reduce the organic concentration in wastewater significantly (Nah *et al.*, 2000).

MBRs have two main configurations. In the first configuration, the recirculated MBR, the membrane module is outside the bioreactor (Figure 3)Figure 3. In this case, inside-out or outside-in membranes can be used. The second configuration is the integrated MBR (Figure 4). An outside-in membrane is submerged in the bioreactor (Buisson *et al.*, 1998; Cicek, 1998) and a negative pressure is created on the permeate side of the membrane that draws the permeate through the membrane while the MLSS is retained on the outer surface. Pressurized membrane modules are also used for integrated MBR operation but their use has been limited. The differences between MBRs and conventional biological treatment systems are as follows (Buisson *et al.*, 1998; Nah *et al.*, 2000; van Dijk and Roncken, 1997):

- high MLSS concentrations in MBRS of up to 35 g/L,
- increased bioreactor temperatures as high as 40°C, resulting from higher biological activities,
- better effluent quality,
- ease of operation,
- reduced HRT and sludge production, and
- enhanced nitrification due to increased SRT and bioreactor temperature.

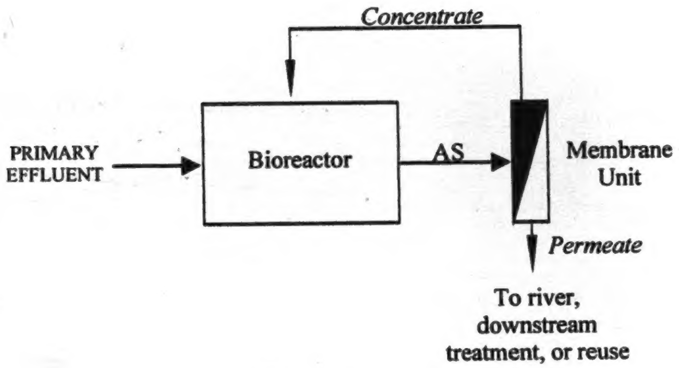
Concentrate
PRIMARY EFFLUENT
Bioreactor
AS
Membrane Unit
Permeate
To river, downstream treatment, or reuse

Figure 3. Recirculated MBR.

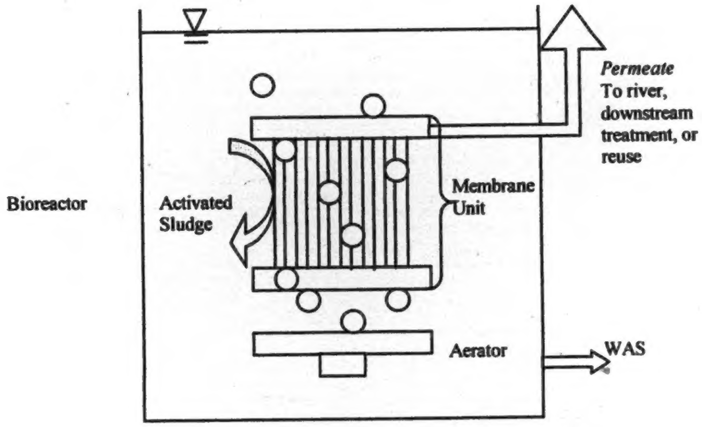
Permeate
To river, downstream treatment, or reuse
Bioreactor
Activated Sludge
Membrane Unit
Aerator
WAS

Figure 4. Integrated or submersed MBR.

The performance and design of the secondary clarifier limits the optimization of the CAS process. MBRs have been developed to ensure treated water is consistently free of suspended solids and a superior microorganism reduction performance is maintained (Buisson *et al.*, 1998). Limitations of MBRs include fouling problems and the subsequent increase of long-term operating costs. Another potential drawback of conventional MBR processes is they may require additional treatment units for removal of nitrogen and phosphorus (Yeom *et al.*, 1999).

2.7.1 Recirculated MBRs

In a recirculated MBR process, low-pressure membranes replace secondary clarifiers for biomass separation. Elimination of secondary clarifiers from the CAS process would reduce land requirements and addresses the concern for bulking sludge and other operational upsets which may affect activated sludge settling in clarifiers.

Many membranes have been studied for the separation of solids from activated sludge. Prior to the recent development of integrated MBRs, high crossflow velocities and transmembrane pressures with either MF or UF modules were thought to be the most effective way to maintain a stable flux and prevent membrane fouling. In the recirculated MBR configuration, activated sludge is pumped directly from the bioreactor to the membrane unit where the solids are separated and returned to the bioreactor. The permeate is further treated or discharged.

The first generation of MBRs utilized UF membranes as the separation mechanism. The ability of UF membranes to completely separate activated sludge from MLSS entirely eliminates the sludge-bulking phenomenon that invariably accompanies the operation of a CAS system (Arika, 1977). UF membranes with a molecular weight cutoff (MWCO) of 20,000 to 50,000 daltons (Da) have commonly been used in recirculated MBRs (Magara and Itoh, 1991).

Arika (1977) investigated the use of UF membranes in the separation of secondary effluent from activated sludge at a 15 m^3/day pilot test facility in Japan. The UF membrane used was an IOPOR unit with a MWCO of 24,000 Da and nominal pore diameter of several nanometres. The HRT was 6 hours. Space requirements for a full-scale facility would be much less than a CAS system do to the elimination of secondary clarifiers. The effluent suspended solids (SS) were consistently less than 2 mg/L and removal efficiencies of 99% were consistently maintained. With BOD loading between 0.03 and 0.2 kg BOD/kg MLVSS day, BOD effluent quality was consistently below 2.0 mg/L with the maximum reported concentration of 4.24 mg/L. Average BOD removal was determined at 99% and COD removal 94%. Problems were not experienced with excess sludge production in the bioreactor even though waste activated sludge was never withdrawn from the system during its seven-month continuous operation. The system was able to maintain a constant membrane flux of approximately 25 L/m^{2}h without once requiring membrane cleaning during the ten-month pilot study.

Chaize and Huyard (1991) investigated the effect of various MBR operating conditions on the removal of organics and nitrogen from activated sludge as well as sludge production. A pilot plant utilizing a DDS UF plate and frame module was used with polysulfone and cellulose membranes. The membranes MWCOs were around 50,000 Da. The transmembrane pressure was in the range of 100 to 200 kPa with a linear velocity of 1.5 m/s. The treated water was of excellent quality with a SRT of 100 days and a HRT of 2 hours. The biomass concentrations varied from 8,000 to 10,000 mg/L with the mass loading of 0.06 to 0.1 kg COD/kg MLSS d. Effluent COD concentrations of less than 30 mg/L were reported with feed concentrations in the range of 250 to 550 mg/L. Influent TKN concentrations of 65 to 150 mg/L were reduced to concentrations of less than 10 mg/L in the effluent.

In their investigation of the use of MBRs for sewage treatment plants in Japan, Magara and Itoh (1991) determined that water temperature and suspended solids concentration should be kept as stable as possible to attain a stable permeation flux in ultrafiltration MBRs. High pressure and high velocity were found to give a high and stable flux; however, the optimum operational conditions in terms of energy consumption and cost were determined to be high pressure and low velocity. The energy costs for pump operation were about 70 to 90% of the total operating costs. They recommended that a membrane filter should be selected on the basis of obtaining a high and stable permeation flux. Polysulfone membranes were shown to have smaller permeation rates than the polyacrylonitrile and polyolefin membranes. In a subsequent study, Magara *et al.* (1992) found that a 20,000 Da UF membrane in a plate and frame configuration could produce a stable flux of 40 to 60 L/m^{2}h at transmembrane pressures in the range of 150 to 300 kPa. Denitrification was possible with the addition of an anoxic tank. Coagulation with alum or ferric chloride was used to remove phosphorus and granular activated carbon (GAC) filters were used to remove organics. Under high loading conditions (1.7 to 2.8 kg BOD/m^{3}d), phosphorus removal was excellent due to the addition of a coagulant; however, the removal of total nitrogen and COD was less than 80% for all the membranes tested.

Ahn and Song Kyung (1999) compared the UF of raw wastewater with the UF of MLSS with an MBR. A UF membrane with a MWCO of 50,000 Da was used in the pilot-study. The transmembrane pressures utilized for the study ranged from 350 to 500 kPa, with a crossflow velocity of 2 m/s. The MBR was more effective than the direct UF of sewage for the enhancement of effluent quality and preventing fouling. In the MBR system, the COD concentration in the effluent was maintained at less than 10 mg/L, regardless of the influent COD concentration. The total Kjeldahl nitrogen (TKN) and total phosphorus (TP) of the effluent were 5 mg/L (63% removal) and 1 mg/L (82% removal), respectively. The MBR was operated for 40 days without cleaning.

Winnen *et al.* (1996) used an inorganic TiO_2 membrane system with a MWCO of 300,000 Da as a recirculated MBR. The membrane was regenerated weekly. The HRT was maintained at about 6 hours. A synthetic wastewater was used to produce mass loading rates of 0.129 kg COD/kg MLVSS d and 0.016 kg TKN/ kg MLVSS d. Sludge production was determined at 0.32 kg TSS/kg COD removed. Removal efficiencies of 95.7 to 99.9% were attained for COD and ammonia. COD in the effluent was 3.1 mg/L with ammonia concentrations less than 0.2 mg/L. An MS-2 challenge test was performed. The initial concentration of the MS-2 solution was determined at 1.5×10^5 plaque forming units (PFU) per milliliter (PFU/mL) using the plaque-forming unit overlay method. At the time of sampling of both the MS-2 feed and the permeate, an MS-2 bacteriophage concentration of 8,200 PFU/mL was observed in the feed solution.

This represented only a 5.5% count of the initial MS-2 concentration at the start of the experiment. The challenge test was repeated with only 3.7% MS-2 count. The low counts may be a result of viruses killed by stress; alternately, the viruses may have been adsorbed onto impurities in the system or onto the membranes. Results of the challenge tests revealed no MS-2 bacteriophage was present in the permeate. Heterotrophic plate counts revealed no presence of bacteria in the permeate as well. Flux levels achieved with this system were not reported.

Cicek (1998) used an inorganic UF membrane (Al_2O_3-TiO_2 monolith) with a molecular weight cut-off of 300,000 Da, for a 500 day pilot study to determine the long-term performance of an MBR system in the reclamation of municipal wastewater. The MBR system had an operational volume of 40 L. A constant HRT of 6 hours corresponded to a membrane flux of 79 L/m^2h, which was maintained throughout the study. A recirculation unit was used with a flow rate of 30 L/min and an average transmembrane pressure of 30 kPa. The membrane was regenerated each week following the manufacturer's recommendations. Nitrate was not effectively removed by this system and effluent nitrate levels were high due to nitrification within the system. All other parameters (i.e. TSS, VSS, COD, dissolved organic carbon, TKN, and TP) were removed with greater than 96% efficiency. Viruses and heterotrophs were not present in the effluent. This indicated that the system was capable of producing very high quality water.

Krauth (1996) used a pressurized UF membrane bioreactor (under 300 kPa) for sludge retention followed by a nanofiltration (NF) unit for additional effluent treatment. The concentrate from the NF unit was fed back to the bioreactor. This enhanced biodegradability and produced a better quality of effluent. The flux through the UF membrane varied from 100 to 200 L/m^2h. Flux through the NF unit was around 0.05 L/m^2h. While attaining a very high biomass concentration (20,000 to 40,000 mg/L) in the bioreactor, the excess sludge production was remarkably small.

2.7.2 Integrated or Submerged MBRs

The first generation of MBRs operated with organic and inorganic membranes placed in an external recirculation loop. The activated sludge was pumped at high speeds (3 to 6 m/s) through the tubes to minimize membrane fouling and reduce the frequency of chemical cleaning. However, the use of recirculation loops leads to increased energy costs (2 to 10 kW/m^3). In addition, the high shear stresses created in the pipes and in the recirculation pumps can contribute to the destruction of bacterial flocs, which has been linked to the loss of biological activity (Seyfried and Brockman, 1995).

The energy costs and the need for extra space for crossflow membrane filtration associated with high speed pumping units lead to the development of submerged or integrated hollow fibre MBRs (Figure 4). With the addition of aeration to the membrane modules, turbulent flow around the membrane surface created reduces fouling while supplying oxygen to the bioreactor (Yeom *et al.*, 1999). Common features of submerged MBRs are:

- direct immersion into the reactor where the biological treatment takes place, and
- operation under an outside-in filtration mode at low transmembrane pressures (less than 50 kPa)

Pore sizes of the integrated MBR membranes are typically between 0.1 and 0.45 µm. Because of this range of pore sizes, integrated MBRs have been typically categorized as MF (Buisson *et al.*, 1998). Integrated MBR systems have been produced which are compact, robust, and cost-effective. Many of the modules can be adapted to replace existing secondary clarifiers. These immersed systems have used either hollow fibre membranes (diameter 0.2 to 2 mm) or flat sheet membranes.

Yamamoto *et al.* (1989) conducted fundamental research in the area of submerged MBRs. His research team was the first to study the effects of submerging a hollow fibre membrane module manufactured by Mitsubishi Rayon Co. into an activated sludge bioreactor. Much work has been done with the Mitsubishi Rayon Co. membrane modules in Japan since. They have been applied to both large-scale municipal wastewater treatment plants as well as smaller scale commercial and residential complexes with a goal of recycling and reusing toilet flushing water.

The first generation of pilot-scale Mitsubishi Rayon MBRs used 0.1 μm hollow fibre membranes. Yamamoto *et al.* (1989) operated these with volumetric organic loads as high as 3 to 4 kg COD/m^{3}d. This corresponded to an MLSS of 30 to 40 g/L. COD and TOC removal were more than 95 and 97%, respectively. Using intermittent suction (10 minutes on, 10 minutes off) at a low-pressure difference of 13 kPa, a stable but very low flux of 0.084 L/m^{2}h was maintained for up to 120 days with an organic loading of 1.5 kg COD/m^{3}d and a HRT of 4 hours. Because of the low suction pressures, energy consumption as low as 0.007 kW/m^3 was achieved.

Recently, the Mitsubishi Rayon unit with the same 0.1 μm pore size was operated at specific flux rates of 10 L/m^{2}h with suction pressures as low as 4 to 6 kPa. The use of intermittent aeration cycles and HRTs ranging from 8 to 15 hours for very high strength domestic wastewaters including toilet-flushing water allowed for excellent nitrification/denitrification systems capable of almost complete removal of ammonia. Yeom *et al.* (1999) used a HRT of 10 hours with an intermittent anoxic and aerated cycle of 60 and 90 minutes, respectively, to achieve an 83% removal of total nitrogen, with almost complete removal of ammonia. Total phosphorus, COD and SS were removed to 55, 97 and 100%, respectively, with the complete removal of coliform bacteria. The unit was operated for 4 to 5 months with no significant accumulation of inert matter and no sludge wasting. Ahn and Song Kyung (1999) observed that, regardless of the variations of influent quality (COD as high as 622 mg/L, BOD at 110 mg/L, TOC at 23 mg/L, SS at 645 mg/L, turbidity at 1 NTU, effluent quality was consistently found to be lower than 30 mg/L of COD, 10 mg/L of BOD, 10 mg/L of TOC, 2 mg/L of SS, and 1 NTU of turbidity with the Mitsubishi Rayon unit.

Using the same unit, Ueda (1996) was able to maintain a flux of 290 L/m^{2}h; however, the reported average power consumption was 2 kW/m^3. Although lower than recirculated MBRs, power consumption was still much higher than CAS processes (Aya, 1994). The membrane was operated without washing for 251 days. It was observed that in the latter stages of operation, sludge viscosity increased considerably after sludge wasting was suspended. This resulted in a substantial suction pressure increase. It was concluded that regular intervals of sludge wasting would alleviate the problems associated with increased sludge viscosity. Intermittent aeration of the bioreactor resulted in total nitrogen removals as high as 83% and the complete removal of ammonia. Total phosphorus concentrations had a 70% removal. TOC, COD, BOD, SS concentrations were all removed with greater than 92% efficiency (Ueda, 1996).

Cheimchaisri (1993) used a Mitsubishi Rayon 0.03 µm pore size membrane for a pilot-scale investigation into the treatment of household wastewater. A stable flux of 100 L/m^{2}h was achieved, with a HRT of 1 day for a low strength wastewater feed with diurnal flow variation. Influent COD concentrations varied from 60 to 200 mg/L, with COD removal efficiency in the range of 80 to 98% for the secondary effluent. Jet aeration improved performance and operation of the MBR.

Using a pilot-scale 0.1 µm hollow fibre membrane module immersed in an activated sludge tank, Visvanathan (1997) determined an optimum air backflushing and filtration cycle. The unit was run with a transmembrane pressure of 13 kPa and a MLSS in the range of 4800 to 5300 mg/L. It was concluded that a cycle of 15 minutes filtration and 15 minutes air backflushing was the most efficient in terms of flux stability and net cumulative permeate volume. More than 90% COD and TKN removal was achieved with the unit.

Another type of integrated MBR is the ZeeWeed® hollow fibre immersed membrane module. This MBR is a proprietary technology developed by Zenon Environmental Inc. Termed the ZenoGem® process, the unit consists of a number of integrated ZeeWeed® membrane modules which form a cassette unit or element that can fit into most existing bioreactors or clarifiers. As pores sizes can range from 0.04 to 0.4 µm, the system is classified in the ultra- to microfiltration range. Other than screening, pretreatment is not a requirement unless phosphorus removal by coagulant addition is desired. Denitrification is an option with the addition of an anoxic tank. A low negative pressure or suction is applied to draw permeate through the pores and into the lumen of the fibres. This has the advantage of low power consumption. Aerators on the bottom of the membrane modules provide turbulence to scour the membranes clean during operation, and also provide the bioreactor with oxygen in conjunction with the bioreactor diffusers. Due to their high chemical resistance, high doses of chlorine could be used to clean and disinfect the modules when flux levels significantly depreciate (ZENON Environmental Inc., 1999). Two hydraulic configurations for the ZenoGem® process can be implemented.

The first configuration allows the membrane unit to be placed in a separate tank from the bioreactor with activated sludge pumped from one tank to the other (Figure 5). This allows for the retrofit of existing CAS treatment systems and also allows the membrane tank to be used as an aerated zone, with part of the bioreactor being converted into an anoxic zone for denitrification. In addition, membrane units can be easily isolated for cleaning and maintenance. The second configuration is simpler with the membrane directly immersed in the bioreactor and a single pump used to extract the permeate under a slight negative pressure (Figure 6). This configuration is typically used for municipal systems with large flows (Husain and Coté, 1999).

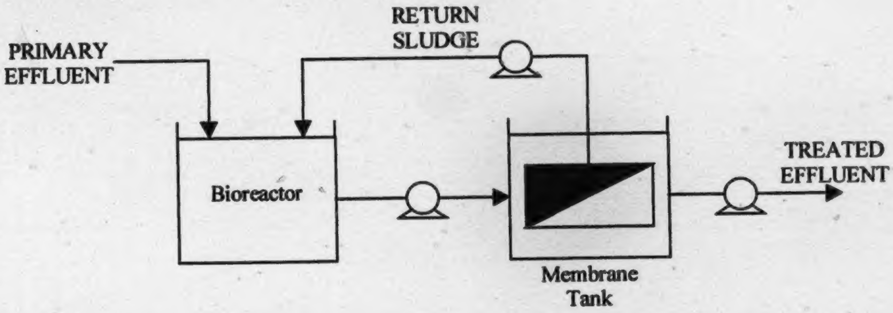

Figure 5. Hollow fibre membranes immersed in external tank after Husain and Coté (1999).

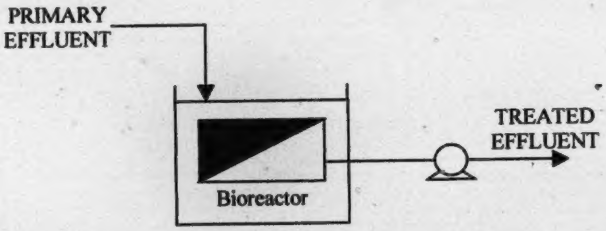

Figure 6. Hollow fibre membranes immersed in bioreactor after Husain and Coté (1999).

In a pilot-scale study with a bundle of hollow fibres (pore size 0.1 μm, filtration area 0.5 m^2) provided by Zenon Environmental Inc., the critical flux was established at 30 L/m^2h. Permeate was withdrawn at a constant rate by suction through the membranes at a fairly low negative transmembrane pressure (less than 40 kPa). During long-term trials, the impact of SRT (10, 20 and 30 days) was investigated in three bioreactors operated in parallel. It was determined that fouling was less severe for an SRT of 30 days. Very high COD removal was obtained (up to 90%) under a high volumetric load of 3.2 kg COD/m^3d and a SRT of 20 days. Sludge production was less for the bioreactor systems than for CAS with only 0.2 kg MLSS/kg COD (Bouhabila *et al.*, 1998).

Operating with a volumetric load of 1.2 kg COD/m^3d and a bioreactor MLSS of 15,000 mg/L, a ZenoGem® integrated UF-MBR achieved a 96% COD reduction and a 95% TKN reduction in the permeate. A 6 log removal of fecal coliforms was also achieved. Sludge production was found to be very low at 0.20 kg MLSS/kg COD. This was due to the long SRT of 50 days in the bioreactor (Buisson *et al.*, 1998; Coté *et al.*, 1998). Using the same design, Coté *et al.* (1998) determined that bacteriophage reduction was 3.8 log units. Ueda and Horan (2000) showed that even with a 0.4 μm pore size polyethylene membrane supplied by Kubota Co., Japan, a 2.3 to 5.9 log reduction of indigenous bacteriophage could be achieved. With this pilot-scale system, a 7 log removal of fecal coliforms was also observed. In the absence of MLSS, removal of phages was negligible, suggesting that phage removal is not a filtration mechanism, but a result of the filtration resistance created by a biofilm on the membrane surface.

Pilot studies utilizing the ZenoGem® system were undertaken at the Milton Wastewater Treatment Plant in the Regional Municipality of Halton, Ontario. The primary effluent of the 2,000 m^3/day facility has concentrations of BOD, suspended solids, TKN, and TP of 100, 100, 30, and 5 mg/L, respectively. Using alum addition, total effluent phosphorus levels have been maintained at an average of 0.05 mg/L while ammonia levels have averaged at about 0.24 mg/L. Turbidity values of less than 0.1 NTU have been maintained in the effluent, indicating very efficient suspended solids removal. A chemical cleaning frequency of every six months has been undertaken which has been successful for the low HRT, high alum dosage mixed liquor. Husain and Coté

(1999) have reported the performance of MBRs in the advanced treatment of municipal wastewater can be expected to provide most municipal effluents with a BOD_5 of less than 2 mg/L, SS of less than 1 mg/L, NH_3N of less than 0.1 mg/L, total nitrogen (TN) (cold climates) of less than 10 mg/L, TN (warm climates) of less than 3 mg/L, TP (chemical treatment) less than 0.1 mg/L, and a fecal coliform concentrations of less than 20 CFU/100 mL without UV microbial reduction.

3 Experimental Design

3.1 Experimental Set-up

Multiple MBR bench-scale units were used to determine the effects of influent flow (Q) and a fixed MLSS concentration on a combined BNR system in terms of phosphorus and nitrogen removal and other effluent quality parameters. Three 80 L BNR-MBR reactors were operated at HRTs of 6, 9, and 12 hours, each with a fixed MLSS concentration of 10,000 mg/L. SRT was determined based on the volume of biomass wasted from the reactors to maintain the MLSS concentration of 10,000 mg/L.

3.2 Effect of HRT on the BNR process

When a reactor volume was set, flow was the determining variable of HRT. HRT can be defined as the theoretical retention time of the fluid elements (soluble constituents) in a continuous-flow stirred tank reactor (CSTR) or ideal plug-flow reactor (PFR). HRT is directly proportional to the reactor volume and inversely proportional to the feed flow (Q) entering the reactor. In an experimental design, influent flow is a relatively easy parameter to control. HRT is defined by Equation 9.

For a combined BNR system, one is interested in the effect of flow (Q) on the performance of the MBR in terms of permeate quality. Does increasing the flow (i.e. decreasing the HRT when the volume of the reactor is fixed) affect the removal of nitrogen and phosphorus, as well as carbonaceous BOD, in a combined nutrient removal system? HRT dictates the amount of time a given soluble substrate is in contact with a population of microorganisms. In terms of full-scale design, it is beneficial to determine the minimum HRT at which an activated process can still perform, because it will ultimately determine a facilities capacity to handle wastewater flow and organic loading.

$$\tau = \frac{V}{Q}$$

Equation 9

where: τ = hydraulic retention time, T

V= reactor volume, L^3

Q= flow, $L^3 T^{-1}$

In the case of a MBR, flow into the system is dependent on the flow of permeate leaving the system. Flow of the permeate leaving the system is defined as flux and is equal to the permeate flow divided by the membrane area. At a fixed membrane area, flux will govern the HRT in a MBR system.

Flux is a function of the pressure difference across the membrane; as one increases the pressure difference across the membrane (transmembrane pressure), flux increases proportionately. In the bench-scale study where the membrane area of the unit is fixed, the maximum flux of the membrane will determine the maximum flow the MBR can handle. In a full-scale design, the flux of the membrane will determine the required area of membrane to treat a maximum flow based on the hydraulics of the wastewater treatment plant and the minimum HRT that can be achieved for the process.

3.3 Hollow Fibre Membranes

Hollow fibre membrane elements typically consist of hundreds to thousands of tiny hollow fibres bundled together and sealed at each end with a resin to form a watertight seal between the permeate and feed streams. Fibres can also be arranged so that both open ends are connected to a permeate collection plate. With this type of configuration, hollow fibre modules have a very high packing density. This provides a large surface area for permeate production; also, the modules can handle high solids loads without plugging. This configuration is very easy to clean by backflushing, using either permeate or creating positive pressure with forced air through the tubes to detach particles deposited on the membrane surface (Aptel and Buckley, 1996). Lumen diameter is typically between 0.5 and 2 mm. Since the homogenous fibres are self-supporting , they are limited by the tensile, compressive and flexure strengths of the porous membrane material. This limits their operating pressure and flux to less than those of spiral wound modules (Paulson, 1995). A comparison of membrane module configurations is outlined in Table 2.

Table 2. Comparison of membrane module configurations after Paulson (1995).

	Module Design			
CHARACTERISTIC	Spiral Wound	Hollow Fibre	Tubular	Plate and Frame
Cost	Low	Low	High	High
Packing Density	High	UF-High RO-Very High	Low	Moderate
Pressure Capability	High	UF-Low RO-High	UF-Low RO-Moderate	High
Membrane Polymer Choices	Many	Few	Few	Many
Fouling Resistances	Fair	UF-Good RO-Poor	Very Good	Fair
Cleanability	Good	UF-Very Good RO-Poor	Very Good	Good

For this study, Zenon Environmental Inc. supplied four hollow fibre membranes elements called ZW-10. Manufactured from a proprietary, chlorine resistant hydrophilic polymer, the hollow fibres are flexible and have excellent tensile strength. These properties allow the hollow fibres to withstand vigorous air scouring and cleaning. The resistance of the membranes to chlorine allows them to be cleaned with a hypochlorite solution to reverse fouling of the membrane. The ZW-10 cleaning protocol is outlined in section 4.2.2. When a suction pressure (typically 10 to 40 kPa) is applied to the open end of the hollow fibre, water is drawn from the outside of the fibre to the inside lumen of the fibre where it collects as permeate Figure 7. The ZeeWeed® hollow fibre membranes are 1.9 mm in diameter and have a nominal pore size of 0.04 μm. Each ZW-10 element (Figure 8a) has a membrane surface area of 0.929 m^2.

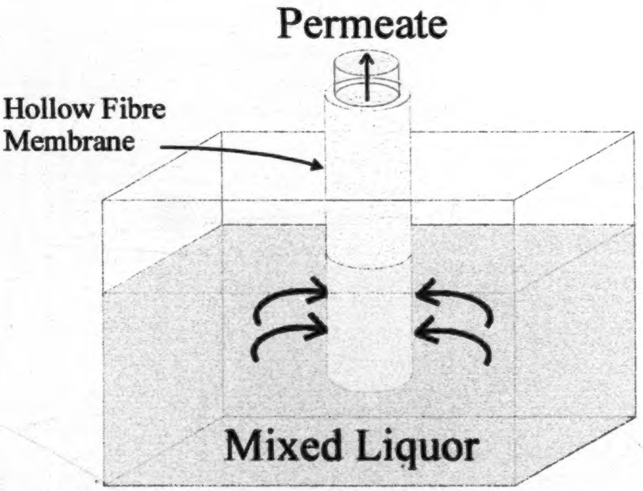

Figure 7. Outside-in flow pattern of hollow fibre membrane.

The hollow fibres are strung between a top and bottom header. The bottom header is a dead end PVC cup. The fibres are embedded into a solid resin to seal the bottom ends of the fibres. The top header contains the loose open fibres sealed within a layer of resin. This vessel collects the permeate from each individual fibre. Separate ports for permeate withdrawal and suction pressure monitoring are located in this portion of the element. An aeration tube runs through the top header to the bottom header of the element forming the backbone support structure of the membrane element. The aeration tube contains small holes near the top and bottom headers from which course bubbles are formed when air is forced through the aeration tube at a flow rate of approximately 28.3 L/min (Figure 8b). Aeration of the element provides O_2 to the activated sludge process in the membrane zone and also provides scour to reduce fouling on the membrane surface.

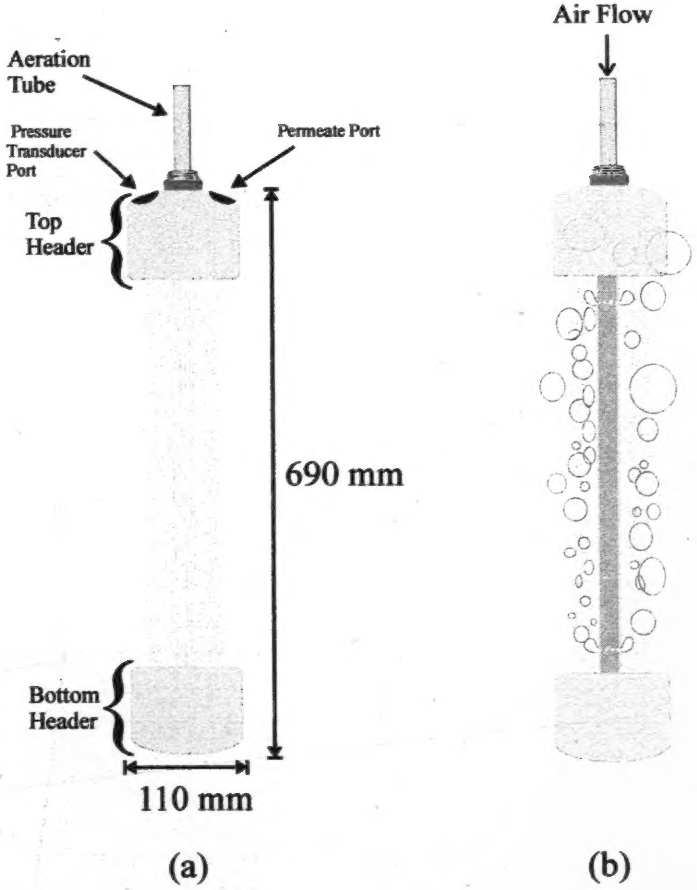
Aeration Tube
Pressure Transducer Port
Permeate Port
Top Header
690 mm
Bottom Header
110 mm
(a)
Air Flow
(b)

Figure 8. (a) ZW-10 element supplied by Zenon Environmental Inc. Membrane surface area is equal to 0.929 m^2. (b) Cross-section of membrane element aeration tube and resulting bubble pattern.

3.4 Equipment Design and Setup

Figure 9 illustrates a conceptual representation of the project configuration.

3.4.1 MBR-Feed System

A continuous primary effluent feed (MBR-Feed) to the three MBRs was achieved by inserting a 250 W submersible pump into the feed channel of bioreactor #1 at GBWWTP. The channel, which was directly connected to the "Y" channel that fed bioreactor #1 prior to its retrofit, provided a continuous gravity feed to the bioreactor from the main primary effluent channel which collects the supernatant from the primary settling tanks of the plant.

From the submersible pump the MBR-Feed was collected into a 0.189 m^3 cylindrical polyethylene tank. A 900 micron mesh strainer attached upstream of the tank prevented large debris (leaves, seeds, etc.) from entering the MBR-Feed system and plugging the feed inlet to the reactor or damaging the membrane element. Constant head in the tank was maintained with a 75 mm overflow near the top of the tank. The overflow was diverted back into the feed channel to bioreactor #1. A 24-hour composite sampler was installed at this location to provide a representative daily sample of the primary effluent entering the MBRs. The sampler was connected to a 12.7 mm stainless steel sampling tube placed inside the tank near the MBR-Feed port. A 25 mm rubber hose connected to the tank was then fed into the sample gallery below the secondary catwalk between the bioreactor and clarifier where the bench-scale MBRs were set-up. This line was connected to a PVC manifold that was reduced to four separate lines; one line feeding each of the MBRs with the other line reserved for temperature monitoring and MBR-Feed grab sampling purposes. Stainless steel fittings were used to unite the PVC lines with flexible 12.7 mm polyethylene lines directly attached to the feed valves of the MBRs.

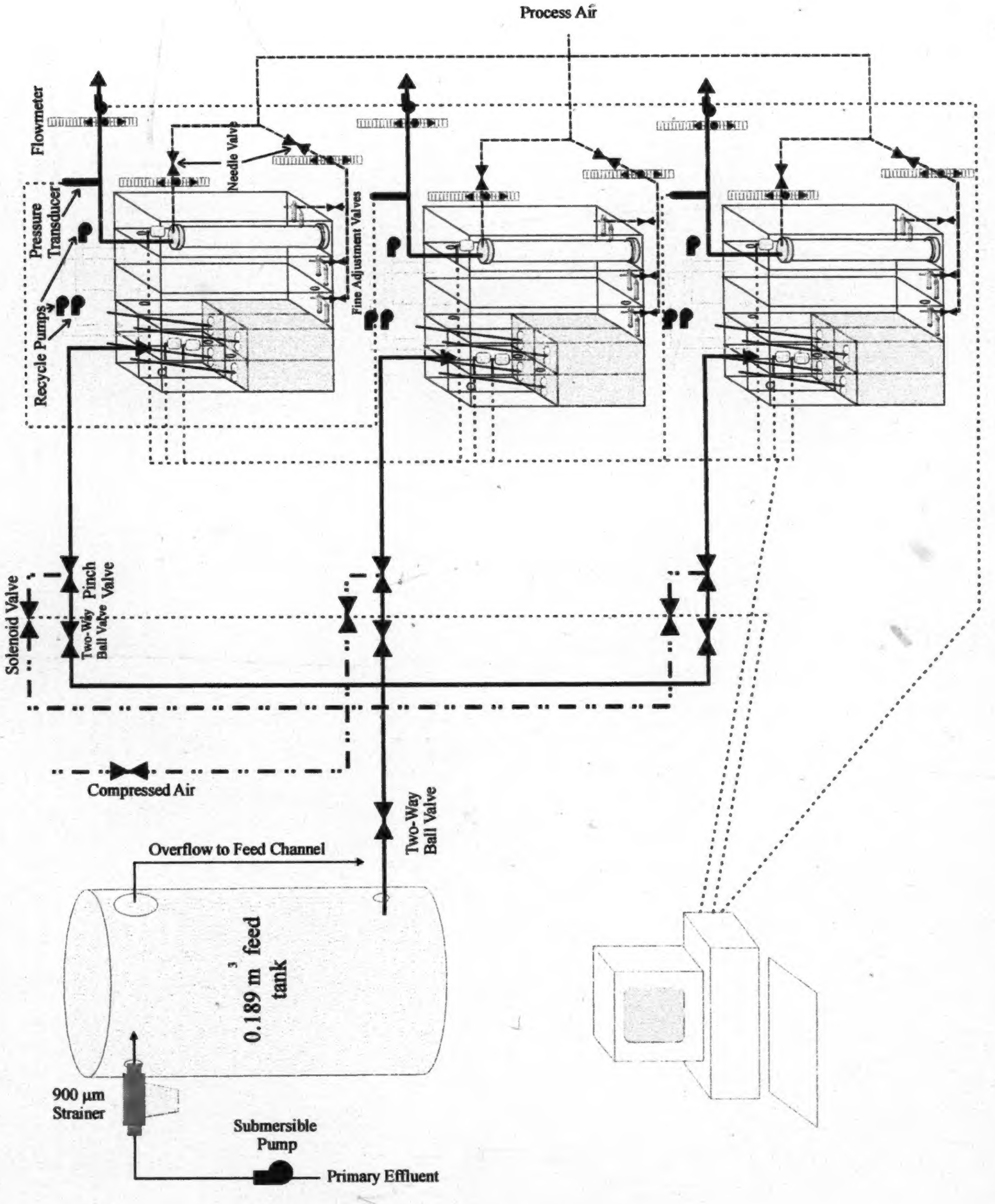
Process Air
Flowmeter
Needle Valve
Pressure Transducer
Recycle Pumps
Fine Adjustment Valves
Solenoid Valve
Two-Way Ball Valve
Pinch Valve
Compressed Air
Overflow to Feed Channel
Two-Way Ball Valve
0.189 m³ feed tank
900 µm Strainer
Submersible Pump
Primary Effluent

Figure 9. MBR project configuration.

3.4.2 Reactor Size

Three 80 L reactors were constructed from 6.4 mm thick plexi-glass. Plan and profile views are illustrated in Figures 10 and 11. The reactors were designed to accommodate the 690 mm tall and 110 mm wide ZW-10 membrane element, with a 150 mm freeboard. Because the reactor dimensions were governed by the ZW-10 membrane element size, conventional plug flow reactor design concepts were not incorporated with respect to length, width and height ratios for this project. Using a design flux of 10 L/m^{2}h, as specified by Zenon Environmental Inc, the reactor volume of 80 L was chosen to produce HRTs of 6, 9, and 12 hours. Substituting volumetric flow Q (calculated as the product of the flux rate times the membrane area) and desired retention time τ into Equation 9, the required reactor volume can easily be determined. A very low design flux rate of 10 L/m^{2}h was chosen to limit fouling during the experimental run so chemical cleaning resulting in reactor downtime could be minimized. A peak flux rate of 30 L/m^{2}h could potentially produce a HRT as low as 3 hours with a reactor volume of 80 L. It was determined that an 80 L reactor would fulfill the theoretical HRT needs as well as allow the membrane elements to be operated within reasonable tolerances with regard to flux rates (Table 3). These operating flux rates would diminish fouling effects during the study and allow continuous operation.

Table 3. Theoretical flux needed to achieve desired HRT.

HRT (hours)	Influent/Permeate Flow (mL/min)	Flux (L/m^{2}h)
6	222	14.4
9	148	9.6
12	111	7.2

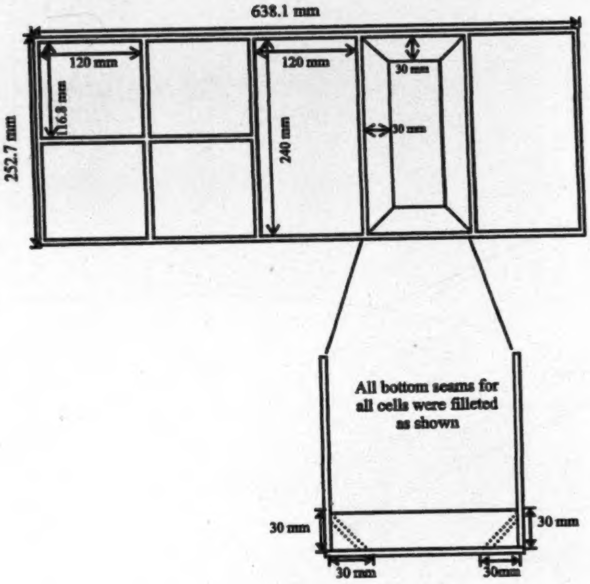
638.1 mm
120 mm
116.8 mm
252.7 mm
120 mm
240 mm
30 mm
30 mm
All bottom seams for all cells were filleted as shown
30 mm
30 mm
30 mm
30mm

Figure 10. Plan view of reactor

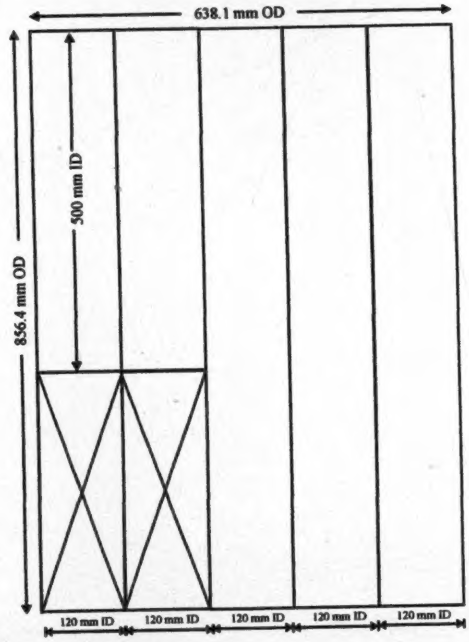
638.1 mm OD
500 mm ID
856.4 mm OD
120 mm ID
120 mm ID
120 mm ID
120 mm ID
120 mm ID

Figure 11. Profile view of reactor. Blocked out portion of reactor was not utilized and only acted as a support structure to the four upper chambers.

Each reactor was designed with seven separate cells: four 5 L cells and three 20 L cells. The reactor was empirically designed into zone volumes to try to emulate the volume ratio of a full-scale 10.57 ML BNR bioreactor at GBWWTP. The ratio of pre-anoxic: anaerobic: anoxic: aerobic used in the design was 1 to 2 to 1 to 12, specifically a 5 L pre-anoxic zone, a 10 L anaerobic zone, and a 5 L second anoxic zone, and a 60 L aerobic zone (Figure 12). The full-scale reactor at GBWWTP utilizes a zone ratio of 1 to 1.2 to 3 to 16. The ratio of the size of the second anoxic zone in the bench-scale reactor was decreased and the anaerobic zone ratio was increased to try to promote the production of VFA within the anaerobic zone through fermentation via a longer SRT than in the full-scale design.

Holes 25 mm in diameter were cut in the walls separating each cell. The holes connecting each cell were alternated up and down and side to side to promote a serpentine flow pattern (Figure 13 and Figure 14). The bottom inside edges of each cell were beveled at 45° to minimize the formation of "dead" mixing zones.

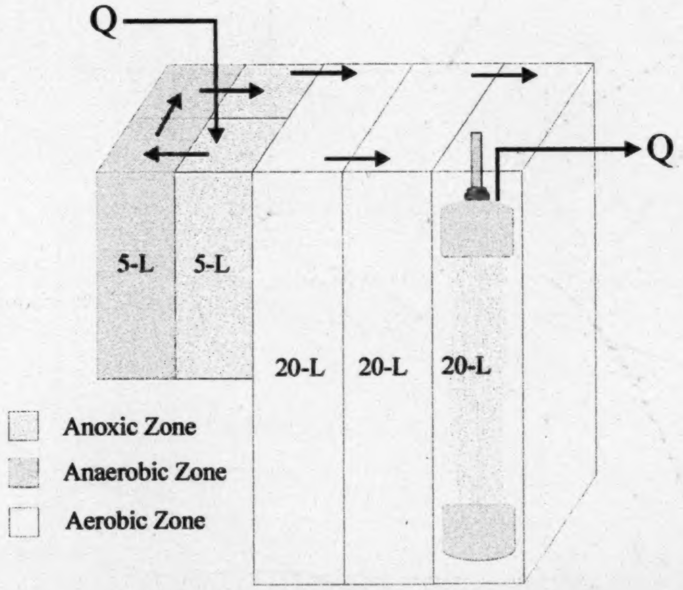

Figure 12. Three dimensional depiction of reactor and corresponding cell volumes.

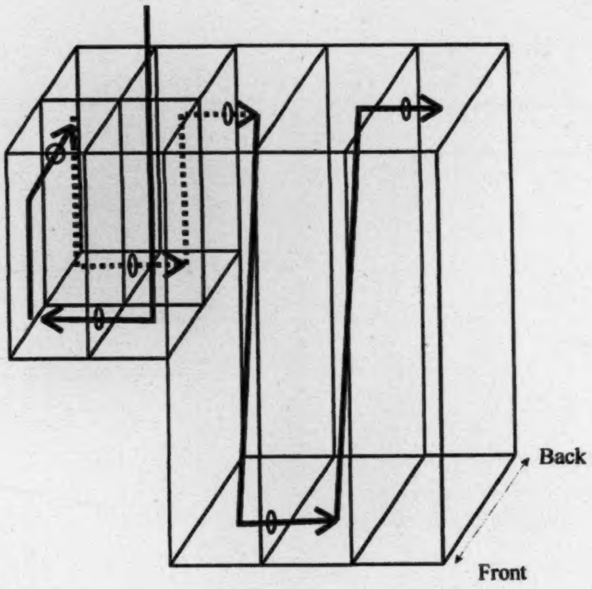

Figure 13. Three-dimensional wireframe reactor illustrating flow pattern.

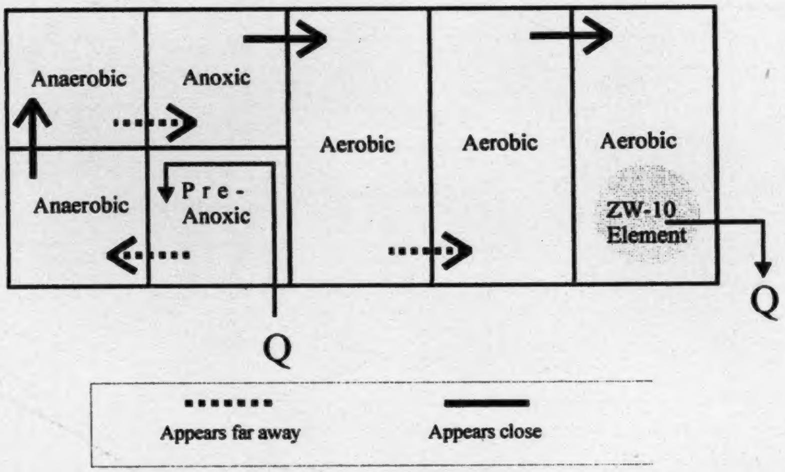
Anaerobic
Anoxic
Aerobic
Aerobic
Aerobic
Anaerobic
Pre-Anoxic
ZW-10 Element
Q
Q
Appears far away
Appears close

Figure 14. Plan view of reactor showing the flow pattern between individual cells.

3.4.3 Liquid Level Control

Liquid level control in the reactors was achieved by utilizing two polypropylene float switches located in the pre-anoxic zone (cell #1) of the reactor. These switches were virtually connected through the computer control software to the influent valve. A pneumatic pinch valve with a natural rubber sleeve (6.4 mm bore) housed within a steel jacket supplied by RedValve® was used to control the influent flow entering the reactor. The valve housing was connected to a "normally open" 120 V alternating current (AC) solenoid valve. Connected to a pressurized air line from a portable compressor, the solenoid valve in the normally open position allowed air to enter the pinch valve and close the 6.4 mm diameter hole. With the solenoid triggered by the float switches, the solenoid closed and released the air pressure through a relief port to allow the rubber sleeve to open. The height of the float switches was set to maintain a volume of 80 L in the reactor with no aeration in the aerobic zones. The switches were mounted to the mixing plate via nylon bulkhead fittings. Table 4 shows the float position and corresponding condition of the solenoid and pinch valves and the influent flow status.

Table 4. Relationship of float position to MBR primary effluent feed system.

Top Float	Bottom Float	Solenoid Valve	Pinch Valve	Influent Flow
Upper Position	Upper Position	Open	Closed	No
Transition	Upper Position	Open	Closed	No
Lower Position	Transition	Open	Closed	No
Lower Position	Lower Position	Closed	Open	Yes
Lower Position	Transition	Closed	Open	Yes
Transition	Upper Position	Closed	Open	Yes
Upper Position	Upper Position	Open	Closed	No

3.4.3.1 Foam Control

During the reactor start-up phase of the project, foaming in the pre-anoxic zone (cell #1) was considerable. This was due to the combination of the low flow recycle entering the zone and denitrification occurring within the zone. The foam build-up in the zone continually caused the liquid level control system to fail due to the build-up of biomass on the two float switches. To remedy this problem, a small sprinkler head was installed on the MBR-Feed inlet to spray primary effluent over the floats. This design change proved successful with the addition of the strainer unit upstream of the MBR-Feed. The strainer unit was beneficial in preventing the sprinkler heads from clogging with debris in the primary effluent.

3.4.4 Membrane Pump

A magnetic drive, 24 V direct current (DC) gear pump was used to generate the suction pressure necessary to produce permeate from the membrane element. The pump speed was manually controlled via a potentiometer and was capable of producing flow rates as high as 2600 mL/min at 1380 kPa. Although more expensive than a peristaltic pump used for sludge recycle, the gear pump minimized pulsation at low operating flow rates (~110 mL/min) and provided more accurate flow control. The pump was originally installed at the bottom PVC header of the membrane element. In this configuration, a siphoning effect was observed when the pump was not operating. In order to achieve complete relaxation of the membrane, where no permeate was produced for a 30 second period, the pump was relocated to the top of the reactor at a higher elevation than the permeate port of the top header of the membrane element. In this configuration, permeate siphoning was not observed during relaxation intervals.

The pump was directly wired to solid-state relays; these relays were then wired to the 24 V DC power supply and to the analog to digital (AD) circuit board in the computer. Via the computer control software (refer to Appendix A), the pump could either be manually controlled "ON/OFF", or be automatically controlled to run a user inputted permeate production interval and relaxation interval. A safety override float switch was also located in the reactor to prevent the membrane element from running dry. When the safety float switch was triggered in the membrane zone, the permeate pump was shut down and put into relaxation mode to allow the tank to refill to the required level.

A flowmeter capable of measuring flow rates between 0 and 600 mL/min was located upstream of the permeate pump. A graduated cylinder and stopwatch were used to confirm the flow rate periodically by measuring the volume of permeate collected in the cylinder in one minute.

3.4.5 Mixed Liquor Recycle

Recycle line inlets and outlets, constructed out of 12.5 mm tubular stainless steel, were placed in the reactor so that recycle flows entering and leaving each cell were near the incoming and outgoing holes of the partitioning walls of the reactor. This was to prevent short-circuiting of the recycle flow. Recycle flow rates were generated with peristaltic pumps and pump heads capable of producing flow rates between 0 and 2200 mL/min. Three peristaltic pumps were required per reactor. The mixed liquor recycles are illustrated in Figure 15. One recycle line (R1) carried mixed liquor from the second aerobic zone back to the pre-anoxic cell operated at 1Q. A second recycle line (R2) carried nitrified mixed liquor from the second aeration zone back to the second anoxic zone. This recycle was operated at flow rates of 1 and 3Q. The final recycle line (R3) carried mixed liquor from the membrane zone to the first aerobic zone. This line was operated at a very high flow rate (5Q) in order to facilitate a homogenous MLSS concentration between the membrane zone and the two upstream aerobic zones. Flow rates for the recycle lines were monitored weekly using "time and fill" method with a graduated cylinder and stopwatch. Mixed liquor from the recycle lines downstream of the peristaltic pumps was collected over a period of one minute. The volume collected over the one-minute period determined the flow rate.

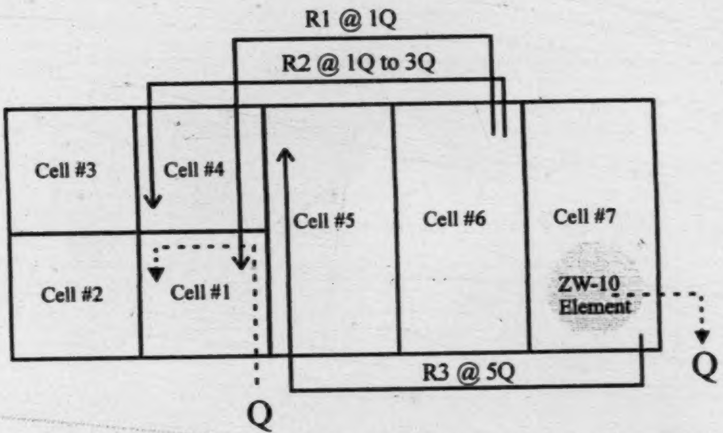

Figure 15. Plan view of MBR mixed liquor recycle configuration.

3.4.6 Aeration

One of the most important aspects of biological wastewater treatment is to provide sufficient aeration for biochemical oxidative processes. In a full BNR system, oxygen is required for the reduction of BOD and ammonia. Aeration in the MBRs was achieved by utilizing neoprene fine bubble diffusers designed for aquariums (Figure 16). These aerators were considered cost-effective, practical, and easy to clean. One aerator was placed in each aerobic zone. The aerator consisted of two 100 mm neoprene tubes coupled to a 6.4 mm vinyl tube housed in a 12.5 mm PVC pipe. This PVC housing was designed to provide rigidity to the aerator assembly and prevent the aerator from moving in the aerobic zone as result of the turbulence from the mixing action of the fine bubbles. The 6.4 mm vinyl air tube was fed through the PVC housing where it was connected to a manifold with fine airflow adjustment valves for each individual aerator. GBWWTP supplied process air for aeration purposes. Two distinct advantages arose from this setup: first, the supply of air was continuous with more than adequate pressure and flow; and second, the air was filtered and dried and of consistent quality.

Airflow was diverted equally to each MBR through a manifold. Prior to the aerators and the membrane element, rotameters capable of measuring airflow between 0.57 and 5.7 m^3/h were installed and frequently monitored. Needle valves downstream of the rotameters were installed to provide fine adjustment of the airflow to both the aerators and membrane element. Aeration was set at a rate to provide a dissolved oxygen concentration of 2.0 mg/L and to provide adequate mixing of the mixed liquor in the aerobic zone. As no mechanical mixers were present in the aerated cells, the aerators were critical to mix the mixed liquor. Placement of the aerators was such that a rolling effect was created to provide turbulent mixing (Figure 17).

An aerator was placed in the membrane zone as well, as the membrane element did not provide thorough mixing in the final zone. In combination with the filleted corners in the reactor design, dead zones were minimized upon visual inspection of the mixing effectiveness of the aerators.

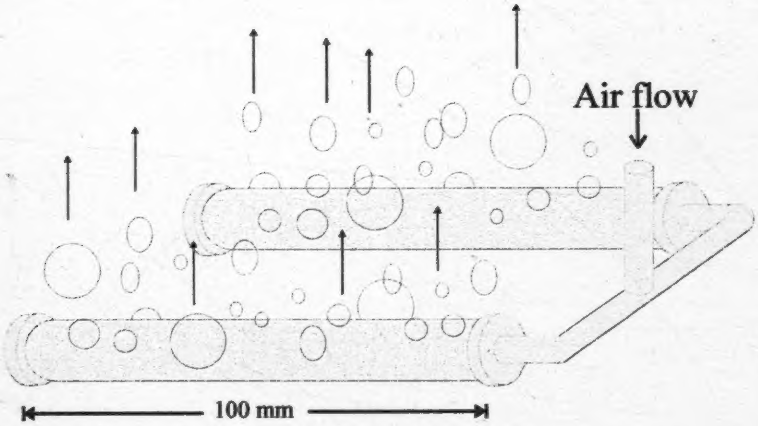

Figure 16. Fine bubble aerator.

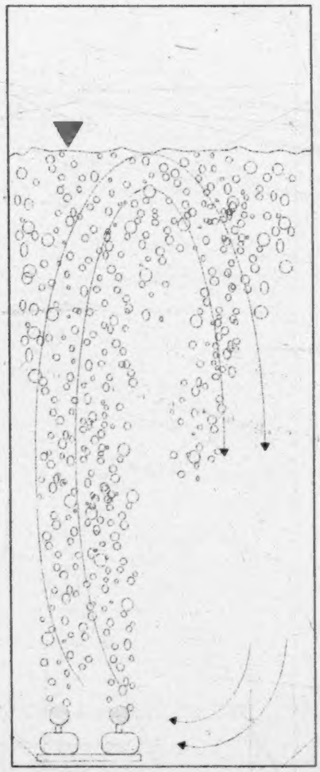

Figure 17. Mixing pattern created by fine bubble aerator.

3.4.7 Mixing

Mixing in the non-aerated cells (cells #1 to #4) was achieved with mechanical mixers. In each cell, an inexpensive 40 rpm, 120 V AC gear motor was coupled to a 340 mm stainless steel shaft with a single blade impeller. Due to mixing inadequacies noticed during the MBR start-up phase, four additional impellers were added to the mixing shafts to increase turbulence in the non-aerated cells (Figure 18). Stainless steel tubing was used to construct the additional impellers. Bottom impellers were constructed by making a sleeve, which slid over top of the existing welded bottom impeller and shaft. By flaring out the tubing, flattening it and providing a 45° pitch to the two blades, the length of the shaft was extended. Similar impellers were manufactured for the middle and top of the mixing shaft.

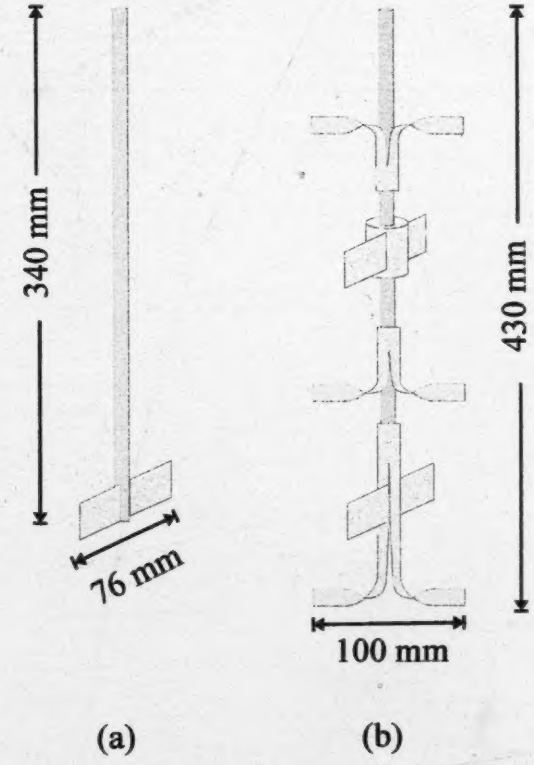

Figure 18. (a) Original mixing shaft and impeller. (b) Modified mixing shaft and impellers.

3.4.8 Online Monitoring

3.4.8.1 Dissolved Oxygen Monitoring

Three analog YSI dissolved oxygen meters were used to monitor the DO in the aerated zones of the MBRs. These meters were connected to the AD circuit board of the control computer. A small 9 V DC amplifier was required to boost the signal from the DO meters to the computer. A correction factor was determined to convert the computer display to match the analog reading on the meter. This correction factor was then inputted into the data logging software to produce the final DO readings.

The dissolved oxygen meters used for online DO monitoring in the MBRs were calibrated using BOD dilution water from the GBWWTP laboratory. The DO of the BOD dilution water was determined using the Winkler's Azide Modification procedure for calibration of the lab meters. A portion of the BOD dilution water was sealed in a 256-mL BOD bottle and taken to the MBR site to calibrate the DO meters. The meters were calibrated weekly and the membranes were replaced when calibrations seemed sluggish and the membranes were visibly fouled.

3.4.8.2 Temperature Monitoring

MBR-Feed temperature was monitored online using a plastic epoxy coated, 25,000 ohm thermoresistor. The thermoresistor was embedded in a 12.5 mm PVC plug that was coupled to the MBR-Feed distribution manifold. A commercial perchloroethylene solvent-based sealant was used to attach the thermoresistor to the plug. The leads from the thermoresistor protruded from the exterior of the threaded plug, where they were connected to the computer through the computer control panel. A 5 V DC voltage was applied to the thermoresistor. Change in temperature would result in a change in voltage from 0 to 5 V. An internal conversion factor was necessary to output the voltage in units of degrees Celsius.

3.4.8.3 Pressure Monitoring

Suction pressure required for permeate production from the membrane element was monitored using pressure transducers supplied by Cole-Parmer. The transducers were directly coupled to the top header of the membrane element via a 200 mm brass nipple.

Requiring a 24 V DC excitation voltage, the pressure transducers were capable of measuring pressures of -101.4 to 206.8 kPa. The transducers were calibrated using a column of water. The calibration curves are presented in Appendix B. Voltage readings were taken at various water level heights in the column. These voltages were plotted against the column water height and extrapolated to the region of negative pressure to produce a linear equation. Equations for each pressure transducer were entered in the computer control software to produce a suction pressure reading displayed in kilopascals.

3.5 Sampling Design

Table 5 outlines the sampling schedule for the study. 8 L permeate grab samples were collected in 10 L polyethylene containers three times per week from a sample port on each individual MBR unit. This sample was split into sub-samples at the GBWWTP laboratory and analyzed for total suspended solids (TSS), 5-day biochemical oxygen demand (BOD_5), carbonaceous BOD_5 ($CBOD_5$), chemical oxygen demand (COD), dissolved organic carbon (DOC), total Kjeldahl nitrogen (TKN), ammonia (NH_3N), total oxidized nitrogen ($NO_2^- + NO_3^-$ = TOXN), total phosphorus (TP) and turbidity. A 200 mL coliform sample was withdrawn randomly for total and fecal coliform analysis. Sterile sample bottles were provided by GBWWTP.

An automated 24 hour composite sampler was configured to collect an MBR-Feed sample three times per week. The sampler was programmed to stop sampling just prior to the collection of the MBR permeate samples. A 200 mL sample was taken every 30 minutes for a total of 48 samples (9.6 L). The MBR-Feed sample was split into sub-samples for analyses at the GBWWTP laboratory. The analytical parameters consisted of TSS, VSS, BOD, COD, DOC, TKN, NH_3N, TOXN, TP, and VFA analysis. VFA analysis was only performed once per week, via a 250 mL grab sample, to match the

GBWWTP schedule. Mixed liquor samples (200 mL) from MBRs 1, 2 and 3 were collected three times per week for MLSS analysis and pH.

3.6 Analytical Methods

All analytical testing was carried out at the GBWWTP laboratory. Analytical methods were based on methods outlined in *Standard Methods for the Examination of Water and Wastewater, 20th Edition* (APHA *et al.*, 1995) and the *Alberta Environment Centre's (AEC) Methods Manual for Chemical Analysis of Water and Wastes* (Dieken *et al.*, 1996). All analytical testing of samples was performed in duplication and, in some cases, triplicate measurements were performed. Quality control samples (standards, blanks, and spikes) were included within analytical sets to ensure quality of data. Sample data validation and acceptance was subject to the quality control guidelines and limits of the GBWWTP laboratory.

3.6.1 TSS determination

The GBWWTP method for total suspended solids determination differs slightly from the method outlined by *Standard Methods* (APHA *et al.*, 1995). A 25 mm fibreglass filter was prepared in a bitumen crucible by washing 10 mL of distilled water three times through the filter and then drying the crucible/filter combination in a muffle furnace at 550°C for 20 minutes. The sample aliquot was directly delivered to the prepared filter/crucible via a Class B wide bore pipette for aliquots less than 100 mL. For aliquots greater than 100 mL a "to deliver" calibrated graduated cylinder was used to deliver the sample. For primary influent a 50 mL sample aliquot was typically used. Because the permeate sample was expected to have a TSS of less than the detection limit of 0.8 mg/L, only 1000 mL of sample were passed through the filter. Although *Standard Methods* (APHA *et al.*, 1995) specifies that a residue of 2.5 to 250 mg must be on the filter to be a valid measurement, this was thought to be impractical for the permeate samples.

After the sample aliquot was delivered through the filter, it was then rinsed with three 10 mL portions of distilled water to wash any dissolved solids through the filter.

The crucible was then allowed to dry in a desiccator for one hour prior to use. The advantage of this method over the Standard Method is the analysts in pre-weighing, filtering, drying, and the re-weighing steps of the procedure do not directly handle the filter. Direct handling of the filter may lead to contamination or damage that may result in false positive or negative results.

3.6.2 VSS determination

The filter/crucible combination used in the TSS analysis was placed in a 550°C muffle furnace for 20 minutes to burn off volatile suspended solids. The crucible was then placed on the counter top for 5 minutes before placement in a desiccator to cool for one hour prior to weighing.

3.6.3 MLSS/MLVSS determination

Although the Standard Method for TSS analysis can be used for the determination of MLSS, the method does limit the sample aliquot that can be filtered through the small fibreglass filter. This not only increases filtration times, but the small sample aliquot may reduce accuracy and precision for the analysis. Gold Bar laboratory uses 95 mm ashless filter (Whatman #1) paper as the filtration medium for the sample aliquot. The weighed oversized filter paper was pressed into a Buchner funnel using an aluminum press. The filter not only covers the perforated bottom of the funnel, but also is creased to partially cover the sidewalls of the funnel. This dish shaped property of the filter paper prevents solids from escaping between the filter paper and the funnel. For the MBR mixed liquor samples, a 25 mL sample aliquot was used. After filtration, the filter paper was folded and dried in 104°C oven for approximately 2 hours. It was then removed and cooled in a desiccator for one hour prior to re-weighing.

The volatile MLSS (MLVSS) were determined by pre-weighing a ceramic crucible, placing the MLSS filter in the crucible, and igniting the volatile solids in a 550°C muffle furnace for 90 minutes. The crucibles were then allowed to cool for 30 minutes, after which, they were re-weighed.

3.6.4 BOD_5 determination

Dissolved oxygen meters utilized in the BOD_5 determination were calibrated using the Winkler's Azide Modification (4500-O C) method outlined in *Standard Methods* (APHA *et al.*, 1995). BOD dilution water was prepared according to *Standard Methods*; however, the water was not aerated prior to use. 25 L of distilled water was dosed with ferric chloride, calcium chloride and magnesium sulfate solutions prior to submersion in a water bath at 20.0 ± 1°C for 24 hours. The phosphate buffer solution was added prior to dissolved oxygen meter calibration.

The BOD procedure follows the BOD_5 procedure (Method 5210B) outlined in *Standard Methods* (APHA *et al.*, 1995) with modifications. Calibrated 256 mL borosilicate glass bottles were used in place of the typical 300 mL BOD dilution bottles. As an alternative to a glass stopper, a plastic cap with a conical liner was used to seal the 256 mL bottle. With the bottle completely filled, the threaded cap displaces a small volume of the prepared BOD dilution water. The cap forms an air/water-tight seal with no air bubbles present.

The prepared BOD sample dilutions were incubated for 120 ± 2 hours in a temperature controlled stainless steel water bath. The dissolved oxygen concentration for all sample dilutions, blanks and references were determined prior to the five-day incubation period. With each BOD set a standard glucose/glutamic acid mixture with a concentration of 20, 200, or 2000 mg/L was used as a quality control measure. This reference was seeded with a stable population of microorganisms to achieve biochemical oxidation of the reference solution. The supernatant of settled raw sewage was used as the seed population for the reference solution and all samples of unknown matrix. MBR samples analyzed for both carbonaceous and total BOD were seeded with 2 mL of the raw influent seed.

Table 5. Sampling schedule. MBR-Feed "Routine" analyses include: TSS, BOD$_5$, COD, soluble COD, DOC, VFA, TKN, NH$_3$N, TOXN, and TP. MBR1, 2, 3 "Routine" analyses include: TSS, BOD$_5$, CBOD$_5$, COD, DOC, TKN, NH$_3$N, TOXN, TP, and Turbidity. "Profile" analyses include: HACH NH$_3$N, NO$_3^-$N, and ortho-phosphate. MLSS sample was analyzed for TSS, VSS and pH.

		MBR-Feed					MBR1					MBR2					MBR3				
Day	Date (2001)	Routine	VFA	Total Coliform	Fecal Coliform	Profile	Routine	Total Coliform	Fecal Coliform	Profile	MLSS	Routine	Total Coliform	Fecal Coliform	Profile	MLSS	Routine	Total Coliform	Fecal Coliform	Profile	MLSS
1	25-Jun	X			X		X		X		X	X		X		X	X		X		X
3	27-Jun	X	X		X		X	X	X		X	X	X	X		X	X	X	X		X
5	29-Jun	X					X				X	X				X	X				X
10	04-Jul	X	X				X				X	X				X	X				X
12	06-Jul	X					X				X	X				X	X				X
15	09-Jul	X					X				X	X				X	X				X
17	11-Jul	X	X				X				X	X				X	X				X
19	13-Jul	X					X				X	X				X	X				X
22	16-Jul	X					X				X	X				X	X				X
24	18-Jul	X	X				X				X	X				X	X				X
26	20-Jul	X					X				X	X				X	X				X
29	23-Jul	X					X				X	X				X	X				X
31	25-Jul	X	X				X				X	X				X	X				X
33	27-Jul	X					X				X	X				X	X				X
35	29-Jul					X				X	X				X	X				X	X
36	30-Jul	X					X				X	X				X	X				X
38	01-Aug	X	X				X				X	X				X	X				X
40	03-Aug	X					X				X	X				X	X				X
45	08-Aug	X	X	X	X		X	X	X		X	X	X	X		X	X	X	X		X
47	10-Aug	X					X				X	X				X	X				X
50	13-Aug	X					X				X	X				X	X				X
52	15-Aug	X	X									X				X	X				X
57	20-Aug	X		X	X							X	X	X		X	X				X
59	22-Aug	X	X				X				X	X				X	X				X
61	24-Aug	X					X				X	X				X	X				X
64	27-Aug	X					X				X	X				X	X				X
66	29-Aug	X	X				X				X	X				X	X				X
68	31-Aug	X					X				X	X				X	X				X
73	05-Sep	X	X				X				X	X				X	X				X
75	07-Sep	X					X				X	X				X	X				X

3.6.5 Ammonia Determination

The GBWWTP laboratory utilizes the automated phenate colorimetric method based upon the AEC method 219 (Dieken *et al.*, 1996) for NH_3N determination and *Standard Methods* (APHA *et al.*, 1995) 4500-NH_3N.

3.6.6 Total Oxidized Nitrogen Determination

The GBWWTP laboratory utilizes the automated cadmium reduction, diazotization colorimetric method for the determination of total dissolved NO_3^- and NO_2^- in wastewater. This method was taken from AEC method 2359 (Dieken *et al.*, 1996).

3.6.7 Total Kjeldahl Nitrogen Determination

The semi-automated block digestion, phenate colorimetric method was used to determine the total Kjeldahl nitrogen in the MBR feed and permeate samples. This method was taken from AEC method 235 (Dieken *et al.*, 1996).

3.6.8 Total Phosphorus Determination

The semi-automated block digestion, ascorbic acid, molybdenum blue colorimetric method was used to determine the total phosphorus content in the MBR feed and permeate samples. This method was taken from AEC method 582 (Dieken *et al.*, 1996).

3.6.9 COD Determination

The GBWWTP laboratory COD determination was based upon the closed reflux, colorimetric HACH method (Method 5220 D). MBR-Feed soluble COD was determined by pre-filtering the sample through a 0.45 μm fibreglass filter.

3.6.10 Dissolved Organic Carbon

Dissolved organic carbon determination was based the AEC method 2411 (Dieken *et al.*, 1996) on the automated phenolphthalein colorimetric method.

3.6.11 Turbidity

Turbidity was determined with the nephelometric HACH method 2130B using a HACH 2100AN turbidimeter.

3.6.12 pH

An Acumet 950 pH meter was used to determine the mixed liquor pH. The meter was calibrated with a three-point calibration using pH buffers 4, 7 and 10.

3.6.13 Coliforms

The method for the determination of total coliforms was based upon the membrane filtration method using mENDO agar (Method 9222B). Fecal coliforms were determined with FC agar (Method 9222D).

3.6.14 Volatile Fatty Acid Determination

Volatile fatty acid concentration (acetic, propionic, butyric) was determined using a gas chromatograph equipped with a flame ionization detector (FID) and Stabilwax-DA megabore column.

3.6.15 Reactor Profile Analyses

For the reactor profile study, ammonia, nitrate, and ortho-phosphate concentrations in the mixed liquor were determined using HACH DR4000 colorimetric methods (Table 6). These methods were found adequate for this purpose because of their ease and quickness at obtaining results. Mixed liquor samples were collected from each zone from MBRs 1, 2 and 3. These samples were then filtered as per the MLSS procedure for suspended solids determination. The filtrate was collected and further filtered through a 0.45 μm cellulose filter to remove any particulate matter that may have passed through the paper filter. Collected filtrate was then used for the HACH analyses.

Table 6. HACH analytical methods used for zone profiling.

Parameter	Description	Range	Detection Limit	HACH Method
Ammonia	Nessler Method	0 to 2.500 mg/L NH_3N	0.017 mg/L NH_3N	Method 8038
Nitrate	Cadmium Reduction Method	0 to 30.0 mg/L NO_3^--N	0.5 mg/L NO_3^--N	Method 8039
Ortho-Phosphate	Ascorbic Acid Method	0 to 2.500 mg/L PO_3^{3-}	0.045 mg/L PO_3^{3-}	Method 8048

4 Results and Discussion

4.1 Reactor Control

Table 7 summarizes the operational conditions for MBRs 1, 2 and 3. The MBRs were operated at theoretical retention times of 12, 9 and 6 hours, respectively. The reactor control results can be summarized in terms of solids maintenance, membrane performance and aeration requirements.

4.1.1 Solids Maintenance

Solids maintenance consists of MLSS control in the bioreactor and SRT. A major operating challenge during the experimental run was to maintain the MLSS concentration in all three MBRs at 10,000 mg/L. MLSS is the major operating parameter in establishing MBR control during the experimental run. As an uncontrolled parameter, SRT was a direct consequence of the volume of sludge wasting to maintain the MLSS in the three reactors.

4.1.1.1 Mixed Liquor Solids Concentration

MLSS and MLVSS were determined over the 75-day study for the three MBRs. MBR 1 was operated at a mean MLSS of 8900 ± 920 mg/L (Figure 19). Due to mechanical failure of the pneumatic feed valve, MBR 1 was offline for sampling days 52 and 57. MBR 1 only reached or exceeded 10,000 mg/L once (day 10) during the experimental run.

MBR 2 was maintained at a MLSS concentration of 9100 ± 2700 mg/L (Figure

20). Sampling day 57 shows a sharp decline in MLSS concentration from 11000 to 2600 mg/L. This was due to an overflow of the reactor mixed liquor activated sludge. The overflow occurred over a three-day period when the unit was left unsupervised.

A MLSS concentration of 11000 ± 1700 mg/L was maintained for MBR 3 during the sampling period. For the period between sampling day 68 (MLSS=11000 mg/L) and day 73 (MLSS=6000 mg/L), a decline in MLSS concentration was observed. Mixed liquor overflow occurred due to extensive foaming in cells 5, 6 and 7 of the reactor. Consequently, a substantial portion of biomass was lost during the overflow event. Microscopic examination and staining techniques (Gram and Neisser stains) of the foam revealed the presence of a large population of *Microthrix parvicella* and *Nocardia* sp. These organisms are common foam-causing agents due to the their hydrophobic cell wall characteristics.

The percentage of MLVSS to MLSS in the bioreactor, sometimes referred to as active biomass, averaged 69.5, 71.2 and 71.7% for MBRs 1, 2 and 3, respectively. This indicates a relatively high amount of inert solids in the bioreactor, a result of low loading and old sludge age experienced with the bioreactors. It was observed during membrane element removal from the bioreactor and subsequent cleanings that large deposits of clay-like material had accumulated on the bottom header of the element in between the individual membrane fibres. The accumulation of this heavy solid matter in the membrane zone of the bioreactors may have affected the solids distribution in the bioreactors resulting in lower observed MLSS concentrations than actually present.

Analysis of variance (ANOVA) was used to determine if the MLSS of three MBRs were in control. A null hypothesis was tested; the three MBRs were operated with the same MLSS concentration over the 75-day period. The alternative hypothesis was that the systems did not have the same mean; the systems were not in control. The ANOVA test revealed the systems were not in control ($F_{Tested} > F_{Critical}$). An ANOVA test was also performed for the operational period between days 1 and 50. During this period the reactors were fairly stable with no overflow events occurring. The ANOVA test revealed the reactors were not in control at 95% confidence. Both ANOVA tests indicate that the effluent quality results cannot be statistically compared. The effluent quality results may be dependent upon the MLSS concentration variability and not the HRT. This supports the conclusion that biological systems are difficult to control when operated in parallel. There are many factors to control in such a dynamic and complex system: aeration, mixing, solids concentration, feed conditions, temperature, recycle flow, permeate flow, etc.

Table 7. Summary of operating conditions for MBR 1, MBR 2 and MBR 3.

	Units	MBR1	MBR2	MBR3
Theoretical HRT	hours	12	9	6
MLSS	mg/L	8900	9100	11000
SRT	Days	242	103	36
Average Organic Loading	$\frac{\text{kg BOD}}{\text{kg MLVSS day}}$	0.056	0.080	0.089
Average Permeate Flow	mL/min	111	150	225
Flux	L/m^{2}h	7.2	9.7	14.5
Average Suction Pressure	kPa	1.5	1.7	4.7
Aeration	m^3/h	2.52	2.35	2.94
Average Dissolved Oxygen	mg/L O_2	4.7	4.5	4.0

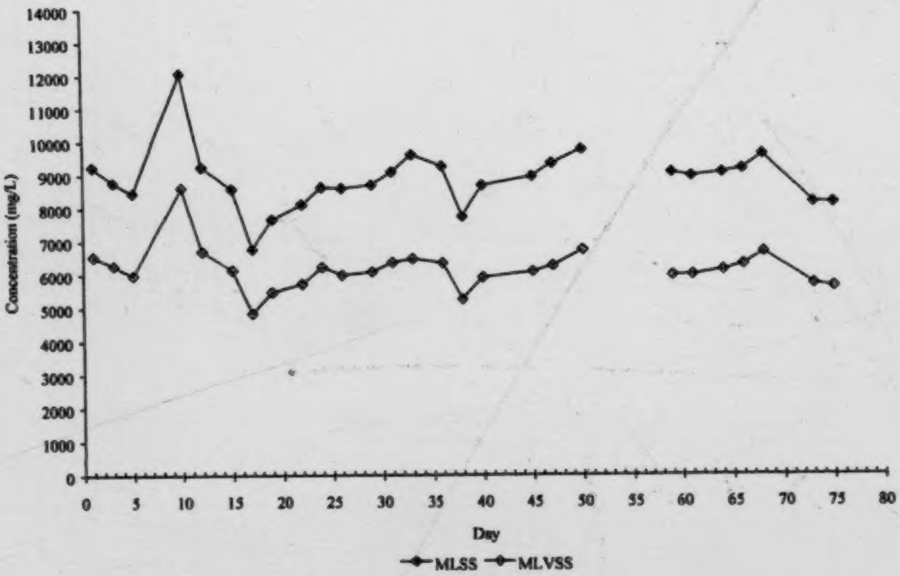

Figure 19. Mixed liquor suspended and volatile solids maintenance for MBR 1.

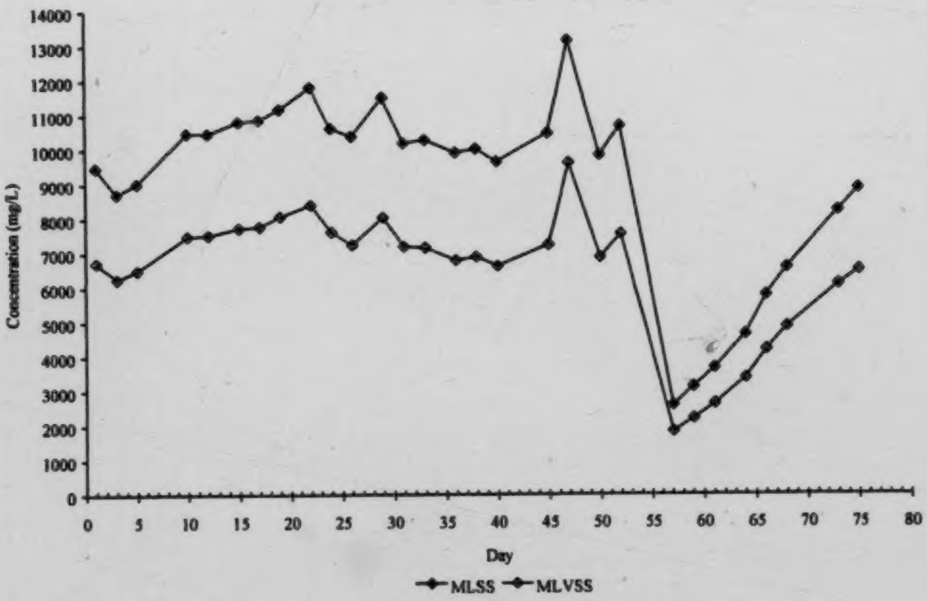

Figure 20. Mixed liquor suspended and volatile solids maintenance for MBR 2.

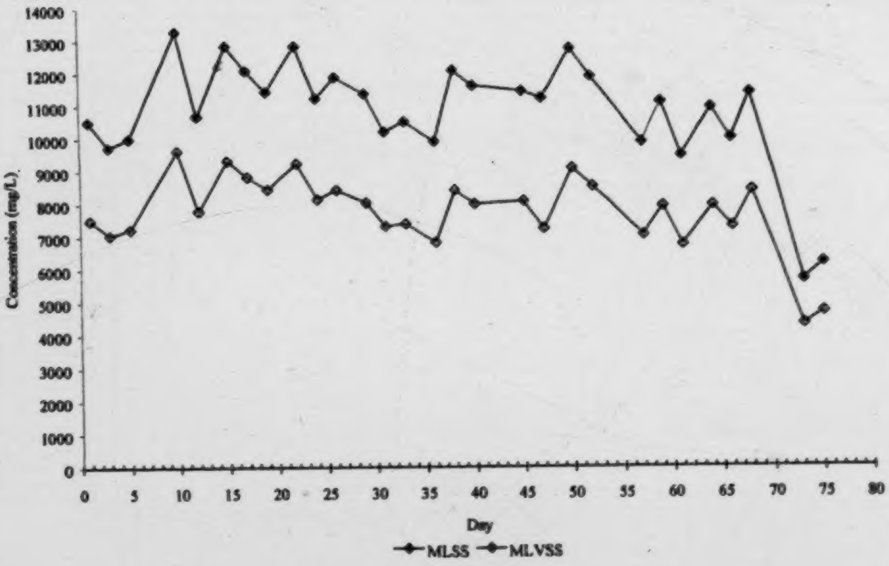

Figure 21. Mixed liquor suspended and volatile solids maintenance for MBR 3.

There was a significant build-up of solids on the sidewalls of the unaerated zones for all three reactors. This was controlled mainly through manual scraping of the biomass from the walls. The scraping may also have effected the solids distribution in the reactors and may have had a negative impact on the MLSS results. Increased mixing intensity in the unaerated zones would have alleviated this problem.

4.1.1.2 SRT

Under the desired operating MLSS concentration (10,000 mg/L) and the flow rates outlined in Table 3, the SRT for MBRs 1, 2 and 3 were 243, 103 and 36 days, respectively. These results are estimates. Due to sludge blanket overflows and foaming issues experienced with the reactors, some biomass removal from the reactors was not accounted for in the SRT determination. Sample removal volumes were accounted for in the SRT determination.

MBR 1 was difficult to maintain near the target MLSS concentration. One explanation for MBR 1 not being able to reach 10,000 mg/L and stay at this level was due to underloading of the reactor. At an HRT of 12 hours, the low feed flow rate into the reactor could not sustain the high growth rate of the bacterial population. Due to the limitation in food supply, the bacterial population entered a phase of declining growth plus the continuing endogenous respiration; the bacterial cells consumed themselves in order to make up for the shortfall in available substrate in the feed. Metcalf and Eddy Inc. (1991) has described this phase of the bacterial growth curve as cryptic growth - a phenomenon in which the nutrients contained in the protoplasm of dead cells, are made available to viable remaining cells as food through cell lysis.

4.1.1.3 pH

The pH was determined on the mixed liquor samples taken from MBRs 1, 2 and 3 for MLSS/MLVSS determination (Figure 22). The pH was fairly stable during the study period ranging from 7.1 to 7.8. This falls within the range necessary for activated sludge viability. There was a slight trend towards an increase in pH; this may be explained due to a net increase in alkalinity due to denitrification.

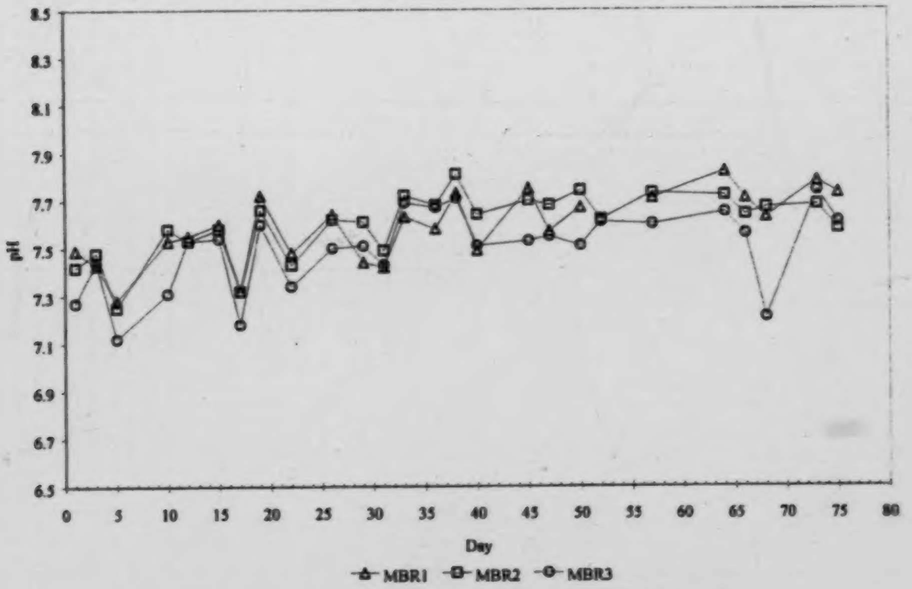

Figure 22. Mixed liquor pH results for MBR 1, MBR 2 and MBR 3.

4.1.1.4 Organic Loading

Commonly referred to as the food-to-mass ratio, the organic loading rates for the reactors (Table 7) are typical for an extended aeration activated sludge process. Table 8 outlines a typical extended aeration activated sludge process.

Table 8. Typical operating conditions for activated sludge systems. Taken from Metcalf and Eddy Inc. (1991).

	SRT (days)	Food to Mass Ratio (kg BOD/kg MLVSS per day)	MLSS (mg/L)	HRT (hours)
Conventional Plug Flow	3 to15	0.2 to 0.5	1000 to 3000	4 to 8
Complete Mix	1 to15	0.2 to 1.0	1000 to 6500	3 to 5
Step-Feed	3 to15	0.2 to 0.5	1500 to 3500	3 to 5
Modified Aeration	0.2 to 0.5	1.5 to 5.0	200 to 1000	1.5 to 3
Extended Aeration	20 to 30	0.05 to 0.15	1500 to 5000	18 to 36

Extended aeration systems are characterized by low loading rates. The active biomass is maintained in the endogenous phase of growth as a result of the low loading rates and long aeration times. The MBRs were operated at a high MLSS concentration compared to an extended aeration system. With no increase in substrate concentration over the conventional full-scale plug-flow system typically operated at an MLSS of 3000 mg/L, low loading (<0.1 kg BOD/kg MLVSS) and high SRT (>30 days) conditions occurred even at an a HRT of 6 hours (see Table 8). MBR 2 showed a higher level of organic loading than expected, primarily due to the overflow event that occurred on day 52 of the study. The significant decrease in MLSS from 11,000 mg/L to 2500 mg/L had a profound effect on the organic loading determination. At the lower MLSS concentration the organic loading ratio was determined to be 0.13 kg BOD/kg MLSS day. Prior to the overflow event the organic loading rate averaged 0.063 kg BOD/kg MLSS day. This result is more in line with the MBR 1 and MBR 3 results.

4.1.2 Membrane Performance

Zenon Environmental Inc. supplied four ZW-10 elements. Each membrane element had a membrane surface area of 0.97 m^2. The membrane elements were individually identified by serial number. Prior to the start-up of the MBRs, the membranes were put through a clean water permeability test after an extended soak (5 hours) in 200 mg/L sodium hypochlorite (NaOCl) solution. Ideally, distilled deionized water would be used to establish baseline permeability; however, high quality water was not available on-site. Potable utility water was used as an alternative. The utility water for the permeability testing had an average temperature of 14.8 to 14.9°C. Over the testing period, an orange color began to appear on the surface of the membranes, indicating the probable removal of ferric oxide from the utility water. Table 9 shows the results of the permeability testing.

Membrane Element W100109 had higher membrane permeability than the other three elements tested. Membrane chemistries may not be consistent between elements; both permeability and nominal pore size may differ between membrane elements as specified by Zenon Environmental Inc. Unfortunately, membrane permeability data for the ZW-10 membrane element cannot be scaled-up due to the differences in the element design and aeration pattern as compared to the elements used for full-scale applications.

Membrane Element	Average Permeability ($L/m^2h/bar$)	Average Temperature (°C)
W100109	855	14.8
W101021	594	14.8
W101028	624	14.9
W101032	607	14.8

Table 9. Clean water membrane permeability results.

Figure 23 illustrates the change in required suction pressure for MBRs 1, 2 and 3 to achieve the flux rates presented in Table 7. Many modifications were required early in the experimental run, which may have affected observed suction pressures. These modifications are outlined in Table 10. Between days 3 and 9, a sharp increase in suction pressure was observed. This can be explained by the relocation of the pumps to an elevation higher than the membrane header and the connection of the pressure transducer directly to the membrane element.

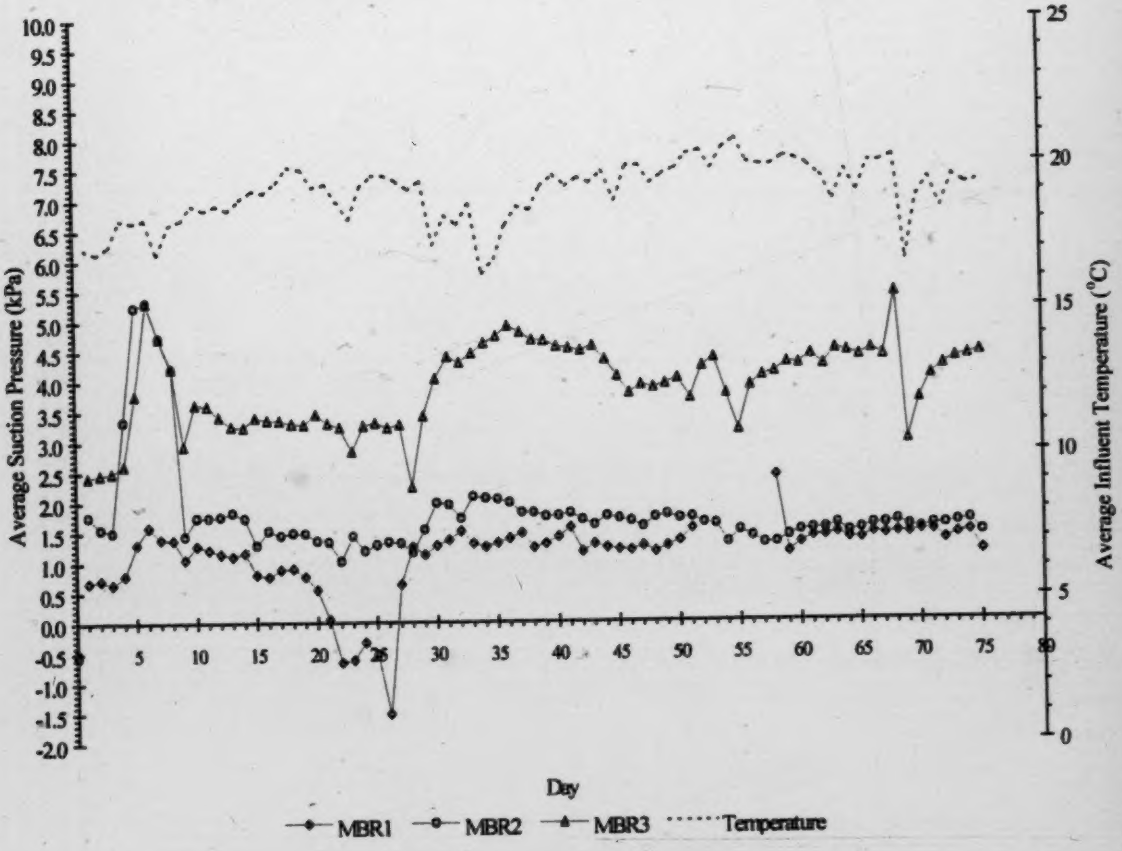

Figure 23. Average suction pressure requirements for MBR 1, MBR 2 and MBR 3.

Table 10. Summary of actions taken that may have affected suction pressure requirements for MBR 1, MBR 2 and MBR 3.

MBR1		
Day	**Membane**	**Action**
1	W101028	Experimental Start-Up
4	W101028	Pump moved above membrane unit
5	W100109	Pressure transducer connected directly to membrane header with extension
23	W100109	Washed biomass from pressure transducer
28	W100109	Second extension added between pressure transducer and membrane header
33	W100109	Tank level very high
44	W100109	*in situ* membrane cleaning
52	offline	
53	offline	
58	W101021	Clean membrane installed
MBR2		
1	W101021	Experimental Start-Up
5	W101021	Pump moved above membrane unit; pressure transducer connected directly to membrane header with extension
23	W101021	Washed biomass from pressure transducer
26	W101021	Second extension added between pressure transducer and membrane header
44	W101021	*in situ* membrane cleaning
53	W101028	Clean membrane installed
MBR3		
1	W101032	Experimental Start-Up
5	W101032	Pump moved above membrane unit; pressure transducer connected directly to membrane header with extension
23	W101032	Washed biomass from pressure transducer
28	W101032	Second extension added between pressure transducer and membrane header
39	W101032	*in situ* membrane cleaning

On day 4, the permeate pump for MBR 2 was moved to a location above the membrane element. This was to ensure the siphoning effect observed with the permeate pump below the membrane element was eliminated during the relaxation interval of the membrane. With the movement of the pump, the pressure transducer location was also changed. Initially the pressure transducer for MBR 2 was below the header level of the membrane element. Because the pressure reading were unexpectedly low, the pressure transducer was moved above the membrane while still attached to the permeate line. This accounts for the high suction pressure reading observed on day 4. With the permeate line elevated, more suction was needed to draw the permeate from the membrane to the permeate pump. On the advice from Zenon Environmental Inc., the pressure transducers for all three MBRs were directly attached to the membrane header of the membrane element via a 100 mm brass nipple. This relocation was to ensure a more accurate reading of the suction pressure. Following day 4, the suction pressures for all three MBRs increased significantly. This was a result of biomass building up on the surface of the pressure transducer. With the high level of aeration in the membrane zone and upstream aerobic zones, a high degree of biomass splashing was observed. Because the pressure transducer relied on a vented cable to produce a pressure reading, the activated sludge blocked the venting port of the cable and resulted in erroneously high suction readings. The biomass was continually washed from the pressure transducers for a period of 28 days until a second extension was added between the pressure transducer and the membrane header. The increased height of the pressure transducer from the water level significantly reduced the amount of biomass splashed on the pressure transducer. Pressure readings began to stabilize around day 33 of the study. The average suction pressures for MBRs 1, 2 and 3 were determined for day 33 to day 75 of the study and are reported in Table 7.

4.1.3 Membrane Fouling

4.1.3.1 Fouling Control

Conventional MBRs utilize backpulsing to minimize the effects of membrane fouling. As fouling increases, as indicated by an increase in suction pressure, permeate production is stopped and a weak NaOCl solution, air or permeate is pulsed from the inside of the membrane lumen to the outside of the hollow fibres at an increased pressure and flow rate. The backpulse pressure removes the biofilm on the surface of the membrane fibres and clears pores plugged by tiny colloidal matter and microorganisms.

Intermittent suction, also termed "relaxation", was used to minimize fouling of the hollow fibre membranes in this study. Between the 20 minute intervals of permeate generation, a 30 second interval was specified in the software to not produce permeate. Aeration of the membrane was continued during the relaxation period to promote the removal of organic and inorganic debris and microorganisms off the surface of the membrane. Lee *et al.* (2001) has shown that a biofilm or dynamic layer of solids and microorganisms is formed on the surface of the membrane when suction is applied across the membrane. In suspended sludge systems, this biofilm or dynamic layer may protect the membrane against fouling from the soluble portion of the mixed liquor. As suction pressure continues, the biofilm or dynamic layer is compacted and results in a decrease in permeate flux. By relaxing the membrane, the layer may be loosened off the membrane and scoured from its surface by the course air bubbles from the membrane element.

Figure 24 illustrates a permeate production run for MBR 3 during an 80 minute period. Intervals #2 and #3 are prime examples of the increase in suction pressure over the 20 minute production interval; interval #2 increases 0.5 kPa and interval #3, 0.4 kPa. Interval #3 shows a recovery of 0.5 kPa after the 30-second relaxation period. Intervals #1 and #2 show a lag period prior to a decrease (0.4 kPa) in suction pressure and subsequent increase during the production interval.

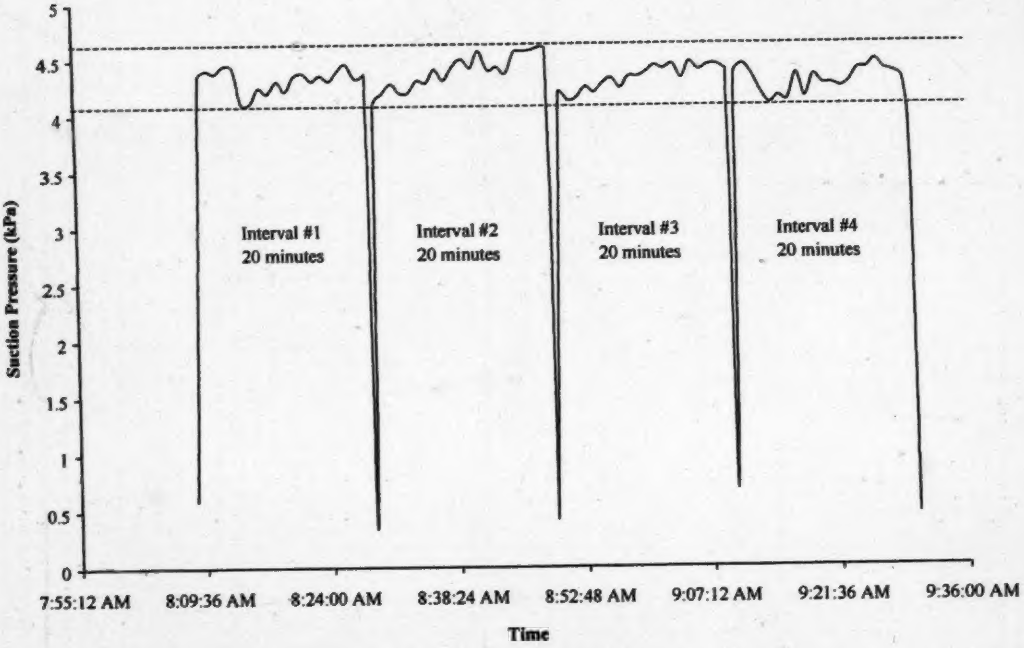

Figure 24. Snapshot of 80 minute permeate production period for MBR 3 on day 59.

4.1.3.2 Chemical Cleaning

The ZW-10 membrane elements did not require chemical cleaning as a result of membrane fouling over the 75-day sampling period. The ZW-10 elements can function at suction pressures up to 50 kPa. However, chemical cleaning was performed on the elements due to presence of microbial contamination within the elements themselves or the permeate pump and lines. Extended soaking of the elements in 200 ppm NaOCl solution did not result in a significant suction pressure recovery. An *in situ* membrane cleaning procedure (refer to section 4.2.2) was performed on the membrane elements on day 39 for MBR 3 and day 44 for MBR 1 and MBR 2. Again, there were no significant improvements in required suction pressure. Due to such low operating flux rates and observed suction pressures, it is very difficult to access membrane cleaning efficiency.

4.1.3.3 Temperature Effects

Due to malfunctions with the online temperature probe, MBR-Feed temperature data was estimated using GBWWTP primary effluent temperature data. MBR-Feed temperature ranged from 16.2 to 20.8°C with an average of 19.0°C. With such a narrow range of feed temperature, membrane performance and required suction pressure were not greatly affected. Figure 23 illustrates there is no apparent correlation between temperature and the required suction pressure.

4.2 Permeate Quality

Table 11 summarizes the permeate quality results achieved by MBRs 1, 2 and 3. Permeate quality is displayed as the mean ± standard deviation and percent removal as compared to the MBR-Feed. Turbidity results are only presented for the MBR permeate.

4.2.1 Solids Removal

TSS removal averaged greater than 99 percent for MBRs 1, 2 and 3 over the 75-day sampling period. The MBR-Feed TSS concentration averaged 91 mg/L. TSS results and permeate turbidity indicate that particulate and colloidal removal from the MBR-Feed is independent of the variation in solids loading to the bioreactors when comparing Figure 25 and Figure 26. Solids removal is directly dependent on the pore size of the membrane. With a nominal membrane pore size of 0.04 µm, the Zenon hollow fibre membranes performed as expected. Turbidity results indicate a high quality of permeate regardless of solids loading on the MBR systems (Figure 27). Turbidity results for MBRs 1, 2 and 3 averaged 0.22, 0.19 and 0.17 NTU, respectively.

Table 11. Permeate quality results for MBR 1, MBR 2, and MBR 3 as compared to the MBR-Feed.

Parameter	Units	MBR-Feed		Membrane Bioreactor 1			Membrane Bioreactor 2			Membrane Bioreactor 3		
		Mean	Standard Deviation	*Mean*	Standard Deviation	Percent Removal	*Mean*	Standard Deviation	Percent Removal	*Mean*	Standard Deviation	Percent Removal
TSS	mg/L	*91*	18	*<0.8*	0.4	99.0	*<0.8*	0.3	99.1	*<0.8*	0.5	99.0
BOD_5	mg/L as O_2	*170*	42	*<2*	0.2	98.6	*<2*	0.3	98.6	*<2*	0.2	98.6
COD	mg/L as O_2	*356*	66	*22*	4	93.3	*22*	3	93.3	*22*	3	93.3
DOC	mg/L as C	*49.9*	14.7	*7.31*	0.85	84.0	*7.39*	0.69	83.3	*7.47*	0.76	83.1
TKN	mg/L as N	*36.7*	8	*0.99*	0.13	97.2	*1.01*	0.24	97.1	*1.01*	0.2	97.1
NH_3N	mg/L as N	*24*	7	*0.07*	0.09	99.6	*0.08*	0.1	99.6	*0.07*	0.09	99.6
Total Nitrogen	mg/L as N	*37.2*	7.2	*12.9*	3.7	65.2	*13*	3.2	64.7	*13.8*	4.5	62.6
TP	mg/L as P	*5.18*	1.31	*3.55*	1.37	42.8	*3.32*	1.33	50.2	*2.86*	2	52.1
Turbidity	NTU	*N/A*	N/A	*0.22*	0.11	N/A	*0.192*	0.09	N/A	*0.17*	0.08	N/A

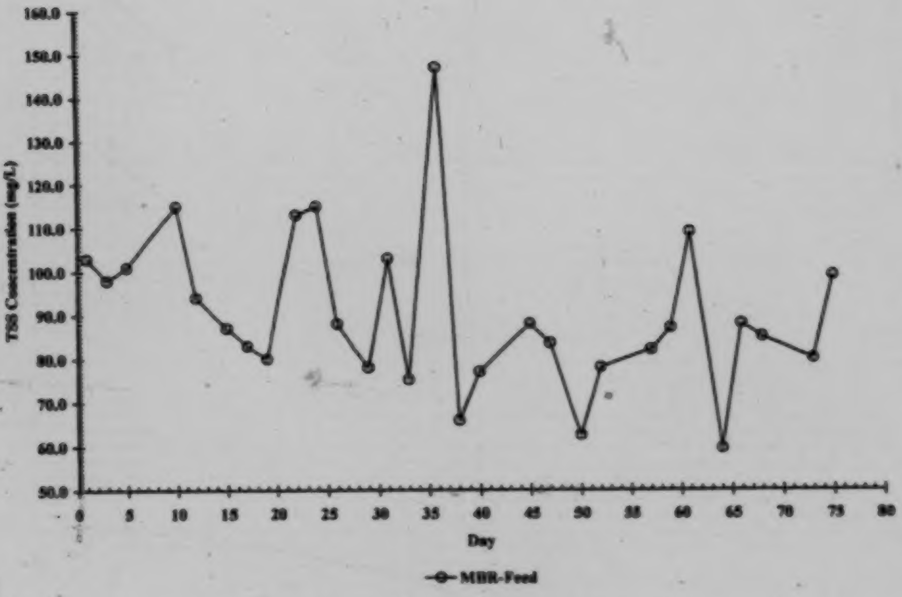

Figure 25. MBR-Feed TSS results.

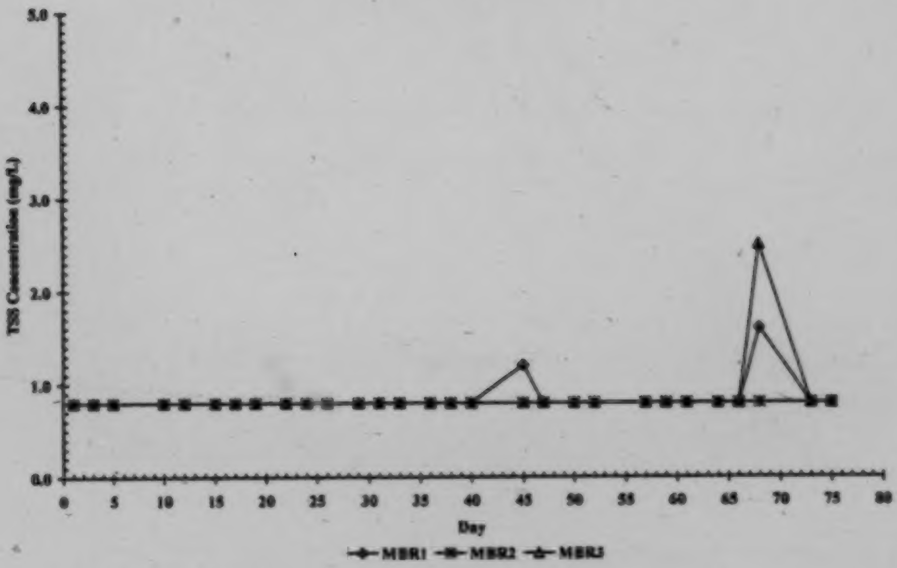

Figure 26. Permeate TSS concentration for MBR 1, MBR 2 and MBR 3.

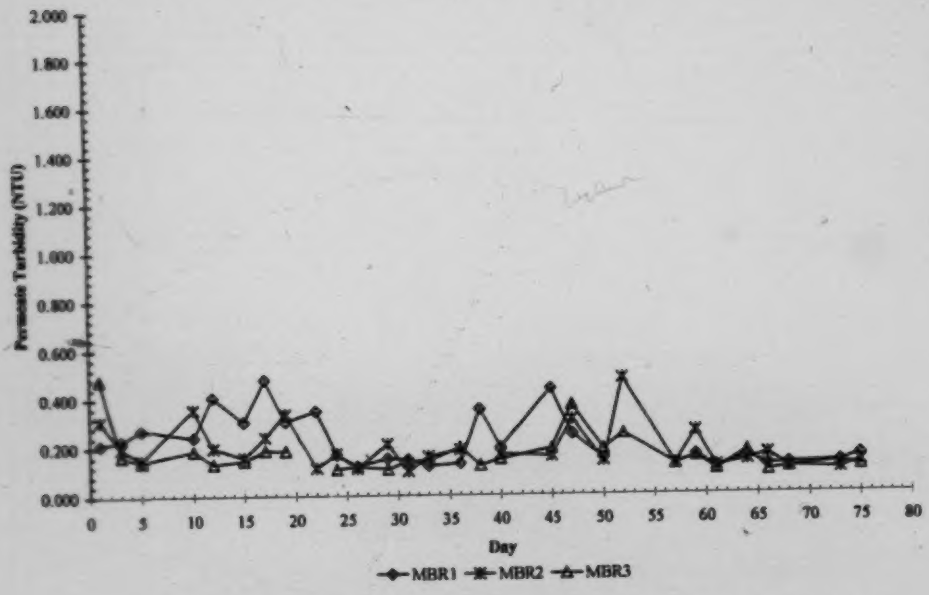

Figure 27. Permeate turbidity results for MBR 1, MBR 2 and MBR 3.

4.2.2 Microorganism Reduction

Coliform samples were collected on three occasions during the 75-day study. On all three occasions, coliforms were present in the permeate samples (Table 12). On day 1, MBR-Feed and permeate samples were only analyzed for fecal coliforms. MBR 1 and MBR 2 had fecal coliform plate counts of "too numerous to count". Because 60 CFU/100 mL is the upper counting limit for fecal coliforms as suggested by *Standard Methods* (APHA *et al.*, 1995), a result of greater than 60 CFU/100 mL was reported. MBR 3 had a fecal coliform result of 1 CFU/100 mL; however, there were a significant number of clear colonies observed on the media plates, possibly indicating a large number of heterotrophic bacteria present in the permeate. MBRs 1, 2 and 3 were sampled on day 3 and analyzed for both total and fecal coliforms. MBR 1 and MBR 2 showed evidence of significant microorganism breakthrough or contamination with greater than 80,000 CFU/100 mL total coliforms and 2,600 CFU/100 mL fecal coliforms in the permeate. This equates to only a 3 log removal of fecal coliforms. Current literature has reported coliform removal as high as 8 log removal and 6 log removal for indigenous bacteriophage from the wastewater stream. Although viruses such as bacteriophage are much smaller than the nominal pore sizes of micro- and ultrafiltration membranes, viruses tend to attach themselves to large particles in the mixed liquor and therefore are effectively removed from the wastewater stream. The buildup of a biofilm layer on the membrane surface may also act as a filtration barrier to which viruses cannot cross. One would expect to observe similar results for this study.

Although excellent suspended solids removal was observed and turbidity levels in the permeate were far below even potable water guidelines, membrane integrity may have been an issue. Suction pressure requirements were quite low for all three MBRs and may also be indicative of a membrane integrity breach. Results seem to indicate contamination within the membrane element or permeate pump and lines downstream of the membrane. Although biofilm is difficult to remove, high concentrations and extended contact times with NaOCl solutions should have eliminated any biofilm on the inside lumen of the hollow fibres or permeate lines. More recent information has led to speculation that poor element construction and quality control, with respect to poor sealing between the permeate header and the aeration tube and/or damaged membrane fibres, may have been the source of the bacterial contamination present in the permeate.

On day 44 an *in situ* cleaning procedure outlined by Zenon Environmental Inc. was performed on all three MBR systems. A 100 ppm NaOCl solution was backpulsed through the membrane element period at a flow rate of approximately 300 mL/min for multiple 30 second pulses followed by 4.5 minute relaxation periods. Following the backpulsing phase, the membranes were aerated for 10 minutes without any permeate production.

Table 12. Total and fecal coliform results (CFU/100 mL) for MBR-Feed, MBR 1, MBR 2 and MBR 3.

	MBR-Feed		MBR1		MBR2		MBR3	
Sample Day	Total	Fecal	Total	Fecal	Total	Fecal	Total	Fecal
1	NA	3.5E+06	NA	>60	NA	>60	NA	1
3	NA	4.3E+06	>8.0E+04	3.9E+03	>8.0E+04	2.6E+03	>100	2
45	3.9E+07	5.5E+06	>160	0	>160	78	>160	0
57	3.9E+07	5.6E+06	NA	NA	>800	74	NA	NA

After the *in situ* clean, total and fecal coliform results from the following day (day 45) still showed a presence of total coliforms for MBRs 1, 2 and 3, although fecal coliforms were not observed for MBRs 1 and 3. Obviously there was still a contamination problem. To determine if the pump and/or the permeate lines were a source of contamination, the permeate lines of MBR 2 were disassembled from the membrane element. A 1000 ppm NaOCl solution was recirculated through the pump, the flowmeter, and all the permeate lines for approximately 20 minutes at the maximum pump output of 2600 mL/min. Total and fecal coliform results indicate coliform contamination was still present and probably a result of the membrane element being contaminated and not the feed lines. One can conclude from these results that once contamination is present within the membrane element or permeate withdrawal system, it is very difficult to revert to sterile conditions. Caution must be exercised when connections are made between the membrane element, pressure transducer and permeate lines.

4.2.3 Carbon Reduction

4.2.3.1 5 Day Biochemical Oxygen Demand

Five-day BOD was reduced to below detection level limits (2 mg/L) for the entire 75-day study (Figure 28). An HRT of 6 to 12 hours provided sufficient detention time for the complete removal of biodegradable organics from an average MBR-Feed BOD_5 of 170 mg/L.

4.2.3.2 Chemical Oxygen Demand

For an average COD concentration in the MBR-Feed of 356 mg/L, MBRs 1, 2 and 3 performed with 93.3 percent removal efficiency (Figure 29 and Figure 30). Average permeate concentrations for MBRs 1, 2 and 3 were 22 mg/L with the very low standard deviations of 4, 3 and 3 mg/L, respectively. COD removal functioned somewhat independently of the MBR-Feed concentration, indicating that COD removal is also a membrane dependent parameter, and not necessarily a process dependent parameter. If sufficient oxygen is supplied, COD removal will proceed to a level in which all the biodegradable compounds are removed metabolically (measured as BOD_5). Remaining non-biodegradable organics are either physically removed through filtration or pass through the membrane pores and present themselves in the membrane permeate.

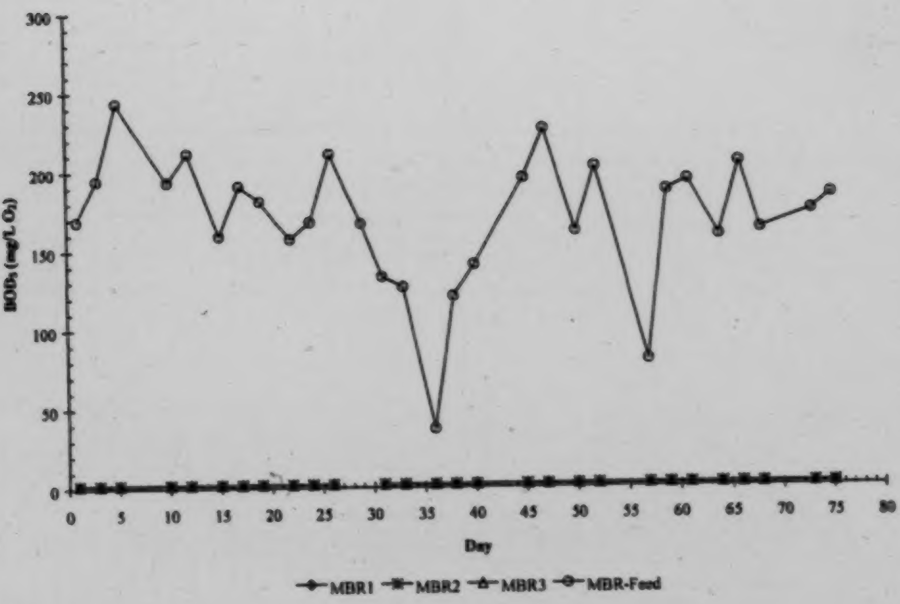

Figure 28. Five-day BOD results for the MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate.

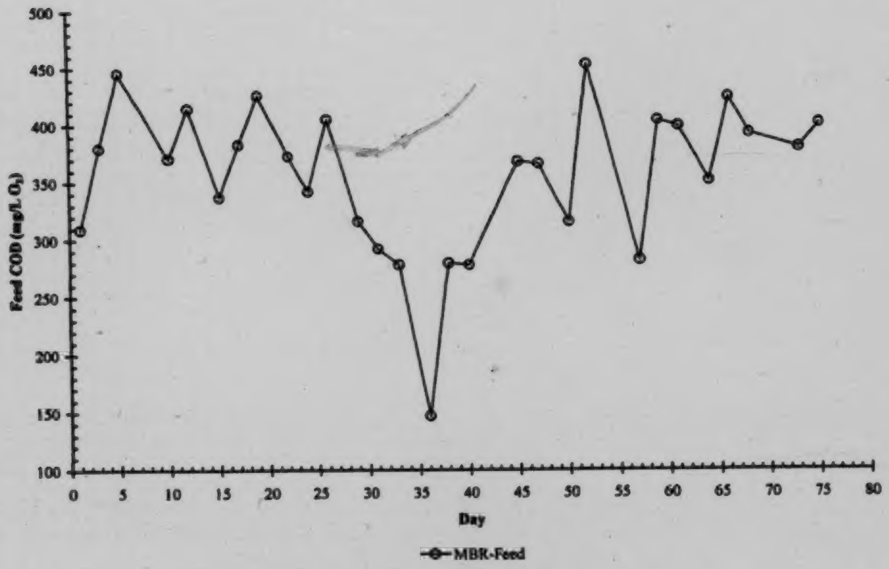

Figure 29. MBR-Feed COD concentration.

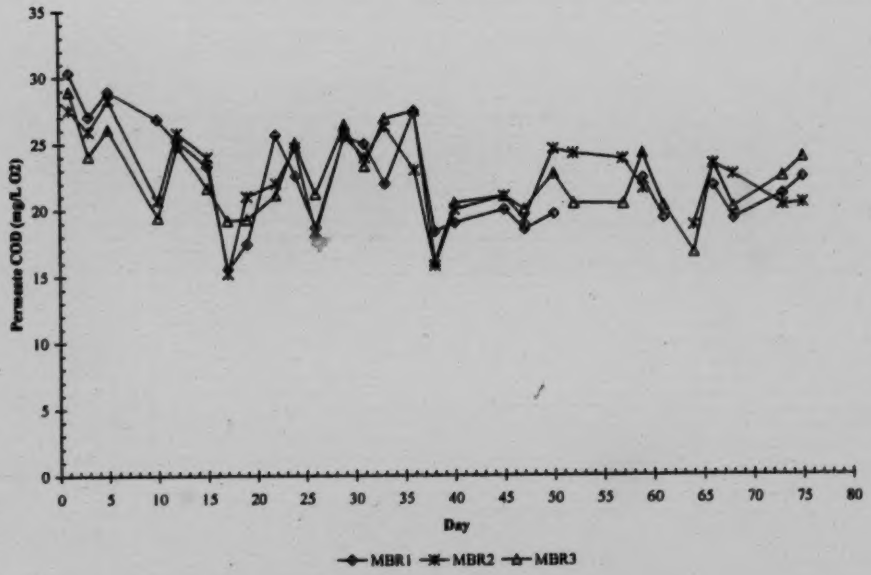

Figure 30. Permeate COD concentration for MBR 1, MBR 2 and MBR 3.

The COD fraction present in the permeate represents those organic molecules that cannot be oxidized metabolically by the mixed liquor activated sludge and are small enough in molecular weight and physical size to pass through the membrane pores. There are two possible explanations for the fluctuation in permeate COD results. The first and the most obvious explanation is the permeate COD fluctuates due to the fluctuation in the concentration of pore-passing organic molecules in the feed. This is supported by the apparent correlation of COD permeate concentration between each of the MBRs as illustrated in Figure 30. Second, the condition of the biofilm layer on the surface of the membrane may change. This layer, which may act as a filtration mechanism for particles and molecules smaller than the nominal pore size of the membrane, may be in a continuous dynamic state. As the layer increases in thickness, smaller molecules may be prevented from passing into the lumen of the membrane, resulting in a lower COD concentration in the permeate. The COD fraction present in the permeate probably consists mainly of aquatic humic substances. These naturally occurring compounds are very difficult to degrade biologically and are responsible for the yellowish hue of treated wastewater effluent. Aquatic humic substances consist of humin, humic and fulvic acids. Humin is not soluble in water and should be removed by the membrane. Humic acid is only insoluble at pH less than two and fulvic acid is soluble in water under any pH conditions. Both these substances would pass through the membrane.

4.2.3.3 Dissolved Organic Carbon

The DOC results supports the COD results presented in the prior section. Regardless of the MBR-Feed DOC concentration, permeate DOC concentration remained consistent between MBRs. With an average feed concentration of 49.9 mg/L, MBRs 1, 2 and 3 permeate averaged 7.31, 7.39 and 7.47 mg/L, respectively (Figure 31). The DOC measured in the permeate represent the fraction of DOC in the feed that cannot be either metabolized by the activated sludge process nor physically removed by the membrane due to properties such as molecular size and weight.

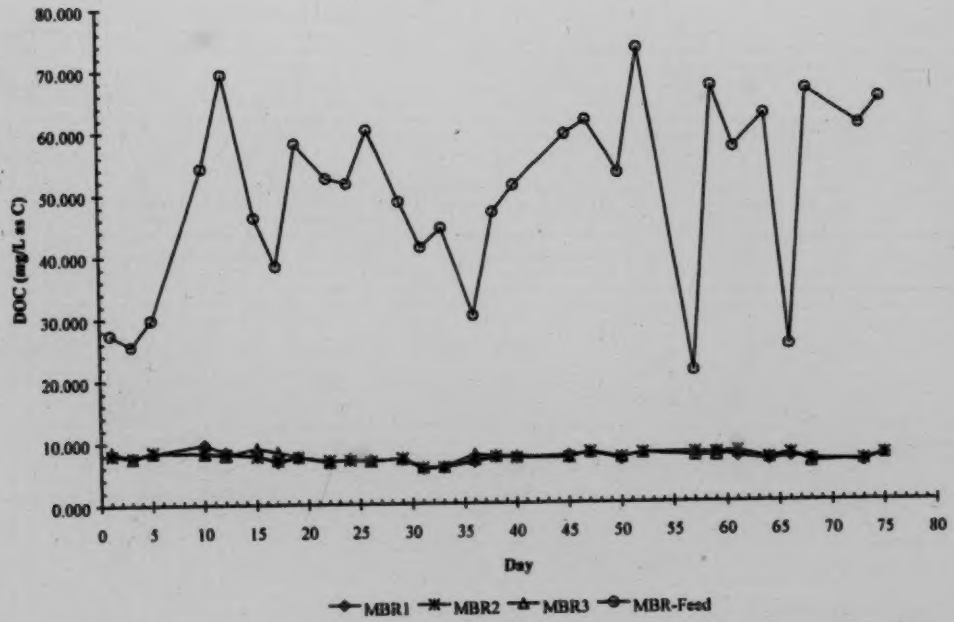

Figure 31. Dissolved organic carbon results for MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate.

4.2.4 Nitrogen Reduction

TKN consists of two components: organic nitrogen and ammonia. TKN was significantly reduced over the 75-day study period, largely due to the complete nitrification of ammonia to nitrate. Ammonia accounts for approximately 64% of the MBR-Feed TKN. With an average feed concentration of 36.7 mg/L, the percent removal for MBRs 1, 2 and 3 was 97.2, 97.1 and 97.1 percent, respectively (Figure 32 and Figure 33). Typically one would expect to see non-detectable levels of TKN in the permeate due to the complete oxidation of ammonia in the bioreactor; however, it was observed that residual TKN levels (mean of 1.0 mg/L as N) were present in the permeate. The TKN is probably in the form of organic nitrogen, a result of the presence of microorganisms in the permeate stream as indicated by the coliform results in section 4.2.2.

All three MBRs had excellent ammonia oxidation explaining the similar performance results with regard to TKN removal. At high SRT and MLSS concentrations, there was little difficulty in maintaining a high percentage of nitrifying organisms within the bioreactors to perform the oxidation of ammonia to nitrate. Ammonia was consistently removed from the MBR-Feed to levels below 0.5 mg/L in the permeate for MBRs 1, 2 and 3, respectively (Figure 34 and Figure 35).

As a consequence to the high levels of nitrification within the reactors, total oxidized nitrogen ($NO_3^- + NO_2^-$ = TOXN) levels were very high in the permeate for all three MBRs (Figure 36). Total oxidized nitrogen levels in the permeate averaged 12.4, 12.5 and 13.0 mg/L for MBRs 1, 2 and 3, respectively. The high TOXN levels in the permeate accounts for the lower than expected total nitrogen removal from the MBR-Feed. High levels of TOXN present in the permeate suggests insufficient denitrification in the anoxic zones of the reactors. In an attempt to increase denitrification in the second anoxic zone (cell #4), recycle flow rates for the recycle between aerobic cell #6 and the second anoxic zone of the reactors were increased from 1Q to 3Q. This increase in recycle rates did not have a significant effect at 95% confidence on total nitrogen removal. A statistical comparison of the three reactors showed no significant difference between the nitrate levels at 1Q recycle in comparison to the nitrates levels in the permeate at 3Q recycle. These results are in agreement with Yoon *et al.* (2000), and supports the theory that nitrate removal may be independent of slight adjustments in recycle flowrates when operating under typical MBR conditions (high MLSS, long SRT).

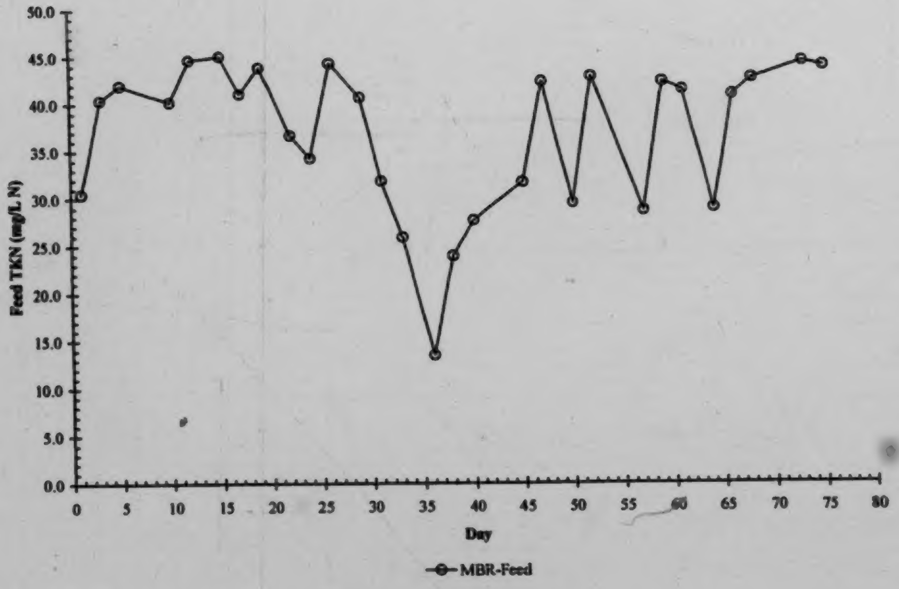

Figure 32. MBR-Feed TKN concentration.

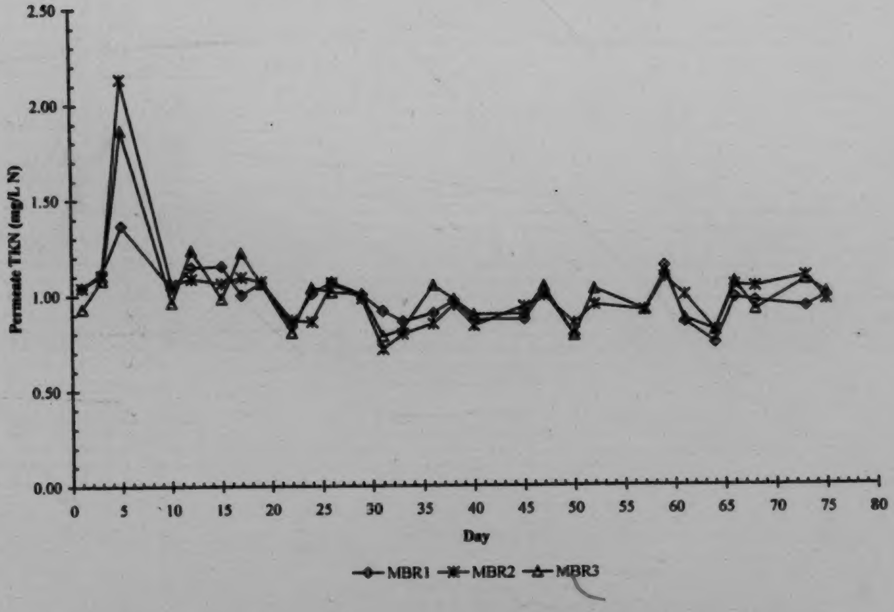

Figure 33. Permeate TKN concentration for MBR 1, MBR 2 and MBR 3.

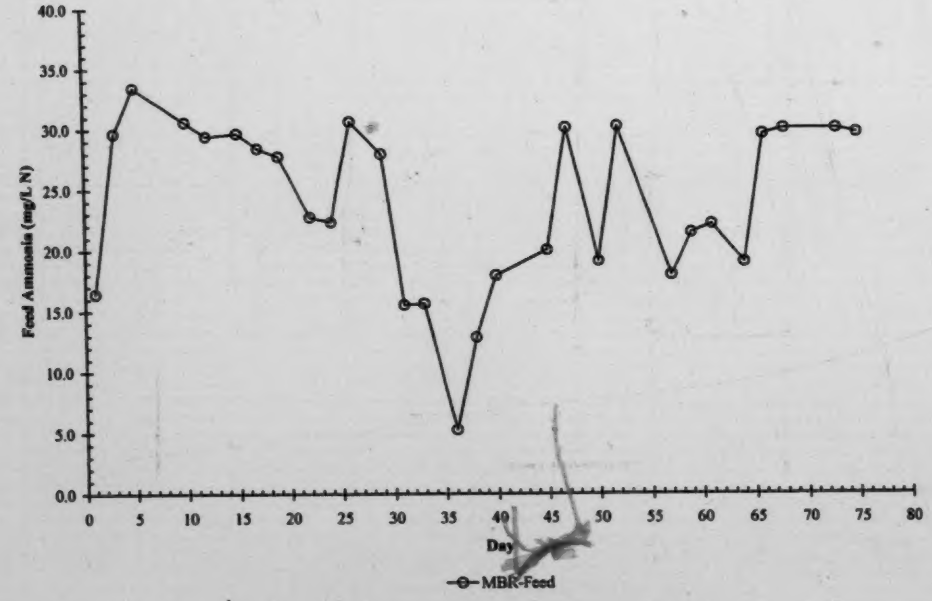

Figure 34. MBR-Feed ammonia concentration.

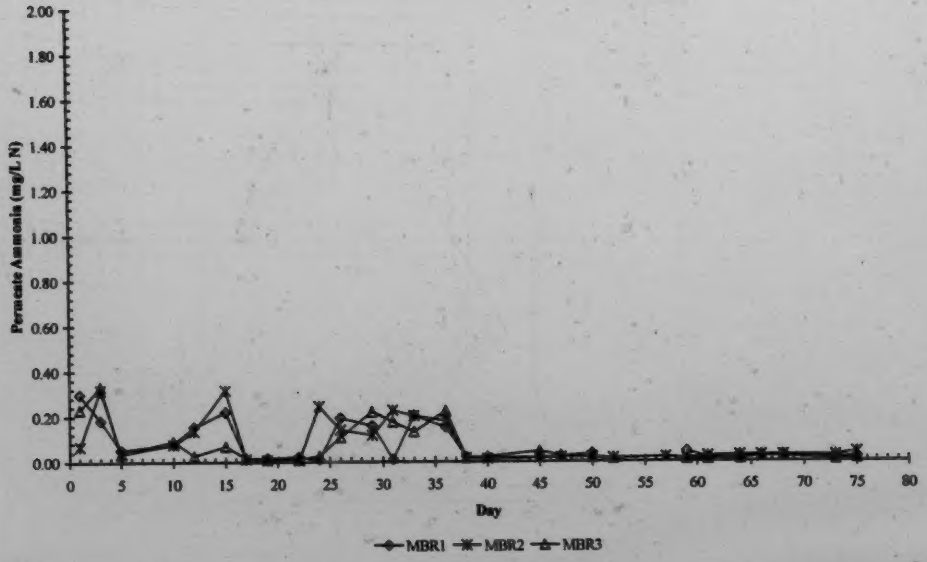

Figure 35. Permeate ammonia concentration for MBR 1, MBR 2 and MBR 3.

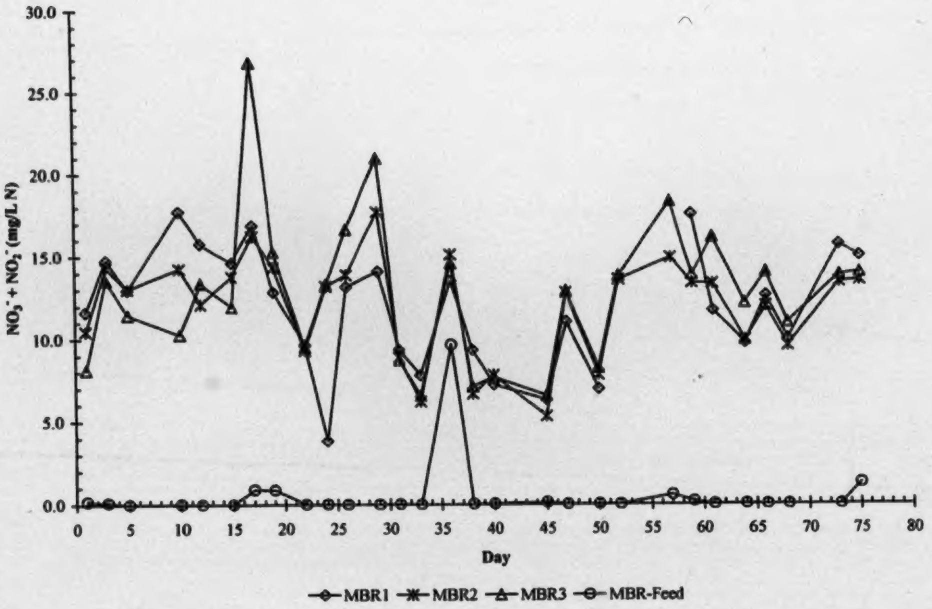

Figure 36. Total oxidized nitrogen (NO$_2^-$+NO$_3^-$ = TOXN) concentration for MBR-Feed, MBR 1, MBR 2 and MBR 3 permeate.

Poor total nitrogen removal, as indicated by the high levels of TOXN present in the permeate, is probably a direct consequence of the MBR design. This is supported by the permeate TOXN results. All three MBRs produced similar TOXN results. It seems as though the second anoxic zone (cell #4) was undersized for nitrate levels present in the recycle mixed liquor. A short HRT in this zone, in combination with the inconsistent levels of easily biodegradable carbon in the influent, would result in insufficient denitrification. A reactor design with a more plug-flow configuration should, hypothetically, alleviate this problem and reduce TOXN levels returning to the pre-anoxic zone upstream of the anaerobic zone. With the recycle of mixed liquor from the aerobic zone (cell #6) of the MBR to the pre-anoxic zone, high concentrations of TOXN (>12 mg/L) were returned to the pre-anoxic zone of the reactor.

Reactor hydraulics dictate that even for an ideal continuous flow stirred tank reactor (CFSTR), 10% of the influent constituents are almost immediately passed through the reactor. The time it takes for 10% of a substrate to pass through the reactor is called the t_{10}, and is an important reactor design parameter. Ideal plug flow reactors have significantly higher t_{10} values and would theoretically allow for more complete denitrification. Because the pre-anoxic zone was one cell with an active volume of 5 L, the pre-anoxic zone, although not ideal due to mixing limitations, can for all purposes be considered a CFSTR. 10% of the returned nitrates in the mixed liquor recycle almost immediately passed through to the first anaerobic cell (cell #2) of the reactor. These nitrates would not have sufficient detention time in the cell for denitrifying organisms to convert them to nitrogen gas. Nitrates passed onto the anaerobic zone are detrimental to phosphorus release and ultimately phosphorus removal in a BNR system. Because the reactor was designed to be a CFSTR in series, it was not possible to simulate ideal plug flow conditions. The t_d for 10% removal of nitrates would be much less for a reactor with CFSTR in series than a plug-flow reactor. This may have significantly affected the denitrification process in the reactor. Further study using a new reactor design with a more plug-flow configuration at the same operating conditions would be needed to test this hypothesis.

In addition to reactor design considerations, high dissolved oxygen levels in the recycle from aerobic cell #6 to the pre-anoxic zone (cell #1) and the second anoxic zone (cell #4) may also explain the poor levels of denitrification within these zones. With the high levels of aeration needed for mixing in the aerobic zones (>4 mg/L D.O.), dissolved oxygen may have been returned to the anoxic zones allowing for the heterotrophic population of denitrifiers to utilize molecular oxygen (O_2) for energy and biomass production rather than nitrate.

4.2.5 Phosphorus removal

With an average MBR-Feed total phosphorus (TP) concentration of 5.18 mg/L, percent removal of phosphorus for MBRs 1, 2 and 3 was 42.8, 50.2 and 52.1%, respectively. Percent phosphorus removal was only determined for those sampling days in which the permeate phosphorus concentration was lower than the MBR-Feed concentration. On numerous occasions, permeate phosphorus concentrations were greater than the MBR-Feed concentration. This phenomenon may have been a result of a secondary phosphorus release in the aerobic zone. MBR 3 produced three sampling days with phosphorus removal less than Gold Bar's total phosphorus license requirement of 1 mg/L. These were days 19 (TP = 0.847 mg/L) and 26 (TP = 0.870 mg/L) with the minimum concentration determined at 0.485 mg/L for day 52. Figure 37 shows a distribution of phosphorus removal for MBRs 1, 2 and 3.

Results presented in Figure 38 suggest that Bio-P removal was effected by the soluble COD concentration in the MBR-Feed. For MBR 3, total phosphorus removal was greatest during periods of high organic loading of soluble COD. On the days where total phosphorus removal was less than 1 mg/L, the soluble COD concentration in the feed was greater than 220 mg/L. For the day with the greatest total phosphorus removal (day 52), the soluble COD concentration in the MBR-Feed was approximately 280 mg/L. This may suggest that TP removal was poor due to insufficient easily biodegradable carbon in the MBR-Feed. This may have effected phosphorus removal in two ways. First, there may have been insufficient easily biodegradable carbon for denitrification in the pre-anoxic zone, which may have resulted in TOXN carry over to the anaerobic zone and therefore inhibited phosphorus release in the anaerobic zone. Second, there was insufficient VFA available for phosphorus release in the anaerobic zone to facilitate phosphorus uptake in the aerobic zone.

The long SRTs observed for the bioreactors may have had a significant impact on the phosphorus removal process. The Bio-P removal process is highly dependent on the physical removal of biomass from the reactor through sludge wasting. Both MBR 1 and MBR 2 had SRTs of greater than 100 days. At this long residence time, in combination with underloaded conditions, bacteria consume themselves through cell lysis. The cytoplasmic contents of the cells are released into the mixed liquor upon cell death, with the macromolecules available to other bacterial cells for cell maintenance and growth. In the case of Bio-P bacteria, intracellularly stored polyphosphate molecules would be released into the mixed liquor and present themselves in the permeate as total phosphorus. MBR 3, which had a shorter SRT at approximately 36 days, produced more acceptable results in terms of total phosphorus removal. This was probably due to the much greater volume of biomass removed through sludge wasting during the study to maintain the fixed MLSS concentration of 10,000 mg/L.

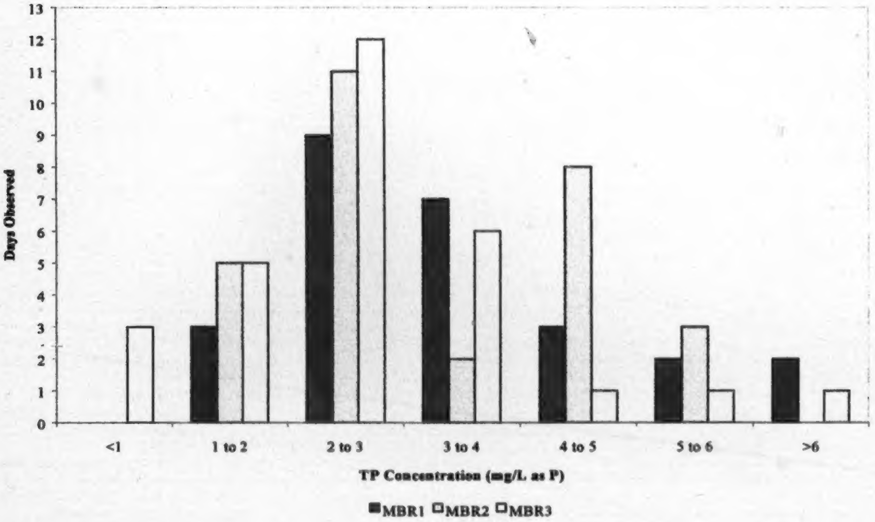

Figure 37. Total phosphorus concentration distribution for MBR 1, MBR 2 and MBR 3 permeate.

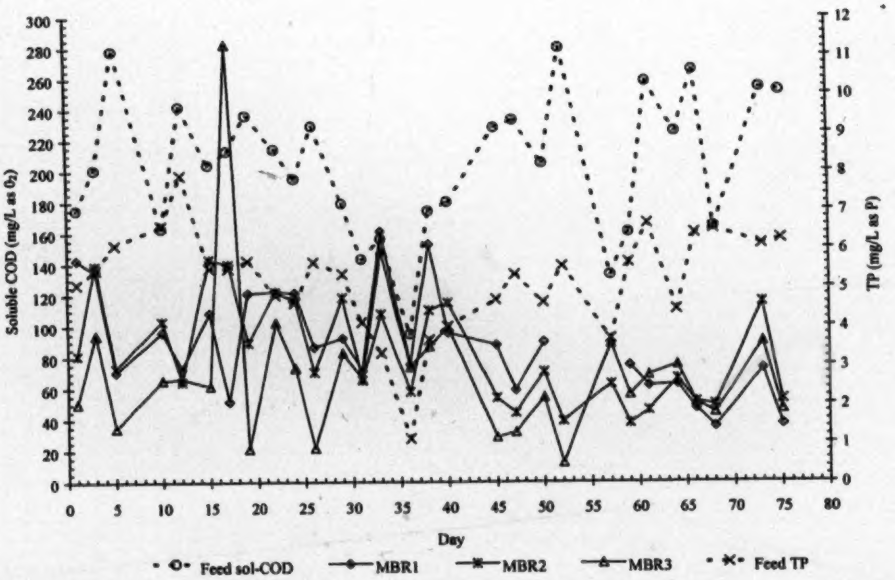

Figure 38. MBR 1, MBR 2, and MBR 3 total phosphorus removal as compared to MBR-Feed soluble COD concentration.

In summary, the poor removal of phosphorus from the three MBRs may be attributed to several factors:

- ☐ The SRT of the reactors was too high for the Bio-P bacteria to be removed from the reactor systems prior to cell lysis and subsequent intracellularly stored polyphosphate molecules were released into the mixed liquor.

- ☐ TOXN produced from high rate nitrification in the aerobic zones of the reactor was carried over into the anaerobic zone from the pre-anoxic zone, essentially reducing the retention time in the anaerobic zone for phosphorus release. This was likely contributed by the underestimation of the second anoxic zone volume required for near complete denitrification. Considered the primary anoxic zone, the short retention time did not sufficiently reduce the TOXN concentration in the mixed liquor returning to the pre-anoxic zone. Due to the reactor design, TOXN in the recycled sludge was very quickly carried over to the anaerobic zone (see section 4.2.4.1).

- There was insufficient easily biodegradable organic carbon in the MBR-Feed, as measured by soluble COD, on a consistent basis to establish steady state conditions and good phosphorus release in the anaerobic zone. Periods of low loading (rain events, weekends and evenings and early mornings) to the reactors, in essence, would uncouple the Bio-P removal process. By the time the process could re-stabilize, the low loading periods would reoccur and the cycle of destabilization/stabilization would repeat itself.

- Over-aeration may have contributed to poor phosphorus removal. During periods of low loading such as after a heavy rainfall event or the weekend, over-aeration may occur whereby Bio-P removal efficiency deteriorates (Doyle and Smith, 1987). A gradual depletion of PHB may occur with the biomass saturated with polyphosphate. With the addition of soluble COD to the system, phosphorus release occurs but the Bio-P bacteria cannot fully uptake the phosphorus in the aerobic zone due to the diminishing amount of intracellularly stored PHB (Brdjanovic *et al.*, 1998a). It is suggested that during periods of low organic loading, aeration should be strictly controlled to prevent deterioration of the Bio-P removal process. For the bench-scale MBR systems, dissolved oxygen levels were consistently greater than 4.0 mg/L in the aerobic zones. Due to the unconventional design of the reactor, aeration was a very important in the mixing process and could not be adequately controlled without sacrificing homogenous mixing of the mixed liquor.

4.2.6 MBR Profiling

A reactor profile was performed on day 35 of the MBR study. Each cell of the reactor was sampled and analyzed for MLSS, ammonia, nitrate, and ortho-phosphate using HACH methods (see section 3.6.15). Unfortunately, on day 35 (July 29, 2002), a significant rain event occurred which resulted in a very low organic loading rate. Volatile fatty acid analysis on the MBR-Feed and first four cells of each bioreactor revealed non-detectable levels of acetic, butyric and propionic acid. Figure 39 shows the profile for MLSS for MBRs 1, 2 and 3. In all three bioreactors, the MLSS concentration averaged between 5,000 and 6,000 mg/L for cells #1 through #3. Biomass increased between cells #3 to #7. Recycle rates were 1Q for the pre-anoxic recycle and 1Q for the second anoxic zone recycle. One would expect little biomass production in the pre-anoxic zone and anaerobic zone as little organic carbon is consumed in these zones for biomass production. MLSS in the pre-anoxic zone is also affected by dilution with the MBR-Feed. In the second anoxic zone where a biomass increase was first observed, dilution does not play a role with the return of mixed liquor from cell #6. This would account for the increase in MLSS in this zone, as mixed liquor with a MLSS concentration of approximately 10,000 mg/L is returned to this zone. As expected, the greatest increase in MLSS concentration occurred in the aerated zone (cells #5, #6 and #7). In cell #5, a significant BOD reduction would result in an increase in biomass.

Because of the heavy rainfall on day 35, the MBR-Feed ammonia concentration was very low at less than 5.5 mg/L. Due to the dilution factor, the ammonia concentration in cell #1 was between 1 and 2 mg/L for all three reactors. A slight increase in ammonia concentration was observed between cells #1 and #4, and may be due to some hydrolysis of organic nitrogen within the anaerobic zone of the reactor. Between cells #5 and #7, ammonia was reduced to less than 0.5 mg/L; it is suspected that ammonia removal was complete.

The Nessler HACH method used for ammonia determination did not produce ammonia values less than 0.3 mg/L for the permeate samples, indicating a slight positive interference during the testing. All permeate samples during the 75-day study were analyzed for ammonia using an autoanalyzer phenate method. These ammonia results showed an average ammonia concentration of less than 0.08 mg/L independent of feed concentration. Permeate ammonia concentrations, such as those observed using the autoanalyzer phenate method, should have been observed for the reactor profile study. The HACH permeate samples were not distilled; this may account for this observed positive interference.

The nitrate profile results are illustrated in Figure 41. MBR-Feed contained an unusually high level of nitrates. This may be explained by the heavy rainfall event.

Nitrates in the raw wastewater entering the GBWWTP may be a result of urban runoff. Commercial fertilizers contain significant concentrations of nitrate. Because of their land application during the summer season, nitrates would be a significant fraction of the storm water runoff, which is partially returned to the wastewater treatment plant via combined storm sewer (CSO) systems.

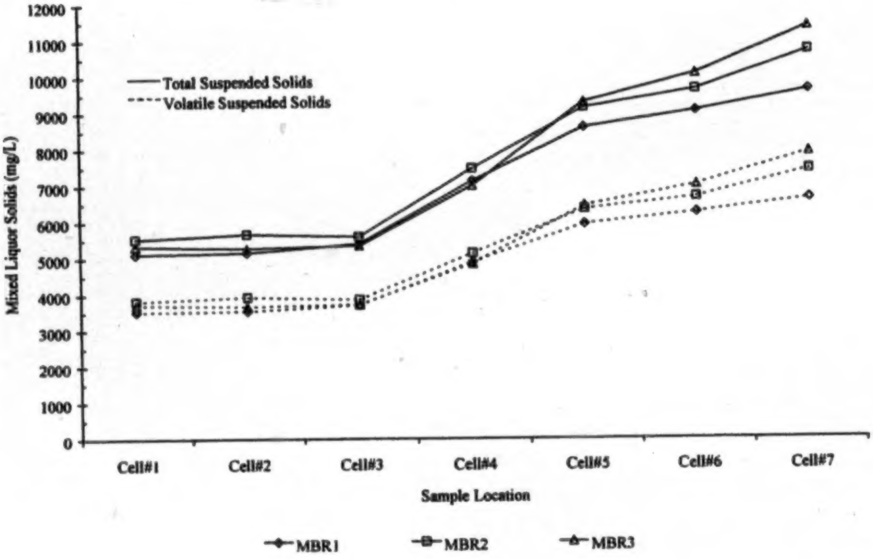

Figure 39. Mixed liquor suspended and volatile solids MBR profile.

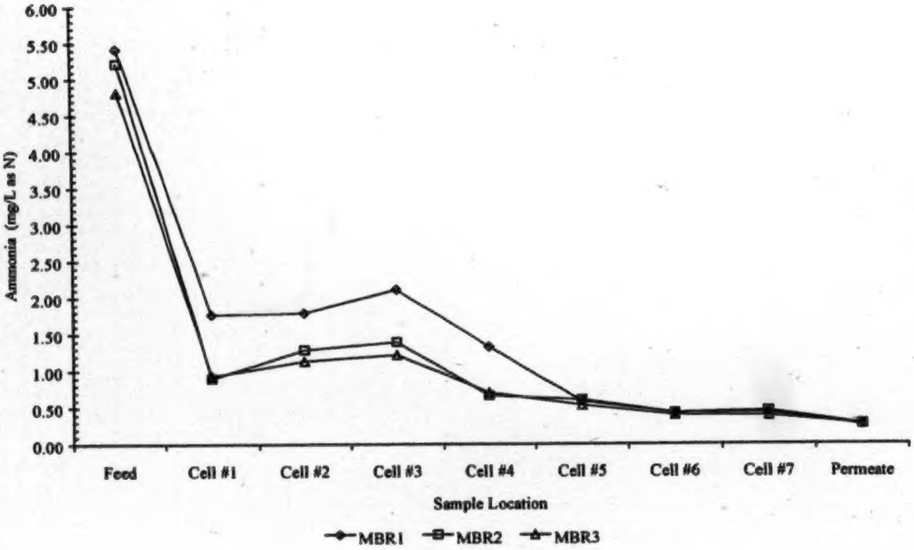

Figure 40. Ammonia MBR profile.

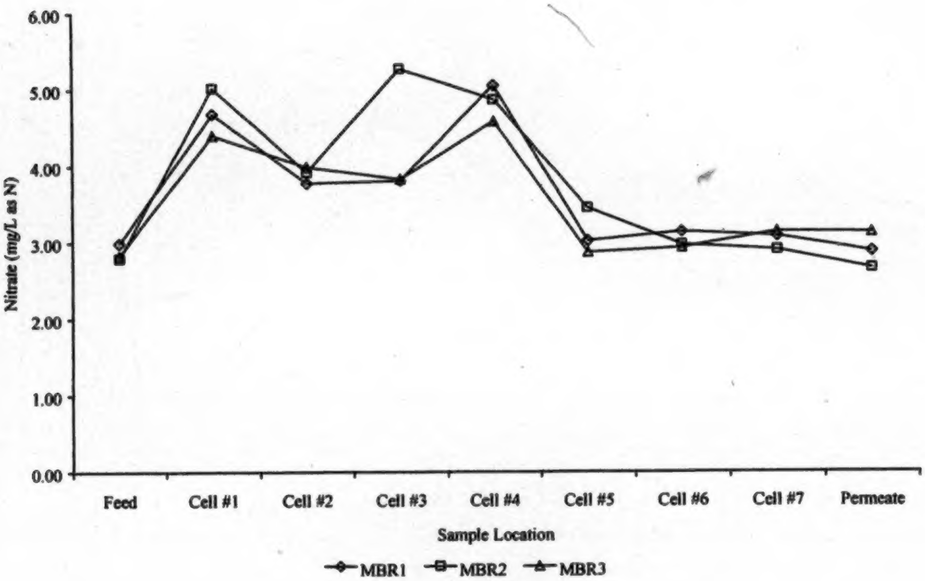

Figure 41. Nitrate MBR profile.

The profile in Figure 41 shows significant nitrate carry-over between cells #1 and #4. This carry over of nitrates would provide strong inhibition of phosphorus release in the anaerobic zone of the reactors. Very little denitrification was observed in the pre-anoxic zone (cell #1) of all three reactors and is probably a combination of two factors. First, the reactor design did not promote a proper environment for good denitrification and nitrate removal. Second, the lack of easily biodegradable carbon in the MBR-Feed due to the heavy rain event would have limited denitrification in the anoxic zones.

The ortho-phosphate removal profile (Figure 42) indicates no Bio-P removal for all three reactors. In fact, permeate ortho-phosphate levels were 2 to 3 times higher than the influent ortho-phosphate concentration. Bio-P removal was completely uncoupled during the profiling study. The high ortho-phosphate levels in the permeate were probably a result of the uncoupling of Bio-P removal in days prior to the profile study. Day 33 showed total phosphorus concentrations of 6.4, 4.3 and 6.0 mg/L for MBRs 1, 2 and 3, respectively.

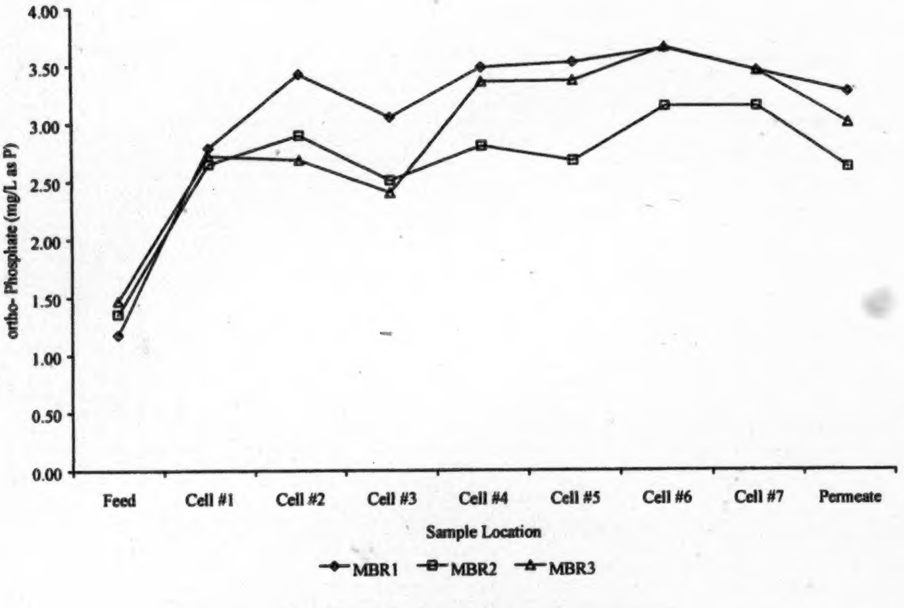

Figure 42. Ortho-phosphate MBR profile.

5 Conclusions

The objective of this study was to compare the performance, in terms of permeate quality, of three MBRs operated at different HRTs. For a full-scale design of a wastewater treatment plant, HRT determines the flow and load capacity and ultimately the size of a reactor to achieve a certain level of effluent quality.

Much research has been performed on MBRs operated in conventional and total nitrogen removal configurations. Little investigation has been conducted into the removal of phosphorus via a chemical free, biological process utilizing MBR technology.

Three bench-scale MBRs were designed, constructed and operated over a 75-day period to determine the removal efficiencies of several common wastewater treatment parameters. For most of the parameters, HRT did not seem to effect removal efficiencies. Membrane dependent parameters such as TSS and turbidity were removed at high levels for all three retention times tested. Complete TSS removal (>99.0%) was achieved as was expected with a membrane of 0.04 µm nominal pore size. Measured turbidity levels in the permeate were consistently below potable water standards, although membrane treated wastewater cannot be considered of potable quality due to the presence of inorganic and organic micro-contaminants. Total and fecal coliforms were present in the permeate, but these results were likely due to a contamination problem downstream of the membrane filtration process and not an integrity issue. Literature and current studies show complete removal of bacteria and protozoa with micro/ultrafiltration membranes, and in some cases, significant removal of surrogate and indigenous viruses.

With respect to the BNR process in an MBR configuration, excellent removal was observed for BOD_5, COD, TKN, and ammonia. Total phosphorus reduction was poor compared to the full-scale process achievements and total nitrogen reduction was lower than expected due to the high level of nitrates in the permeate stream. Total phosphorus levels in the permeate for a 12, 9 and 6 hour HRT were 3.55, 3.32 and 2.86 mg/L respectively, for a percent removal of total phosphorus from the MBR-Feed of 42.8, 50.2, and 52.1 percent. Poor total phosphorus reduction can be explained by a combination of three main factors: a long SRT, which resulted in a secondary phosphorus release in the aerobic zone; TOXN carry-over into the anaerobic zone due to poor reactor design essentially decreasing the anaerobic HRT; and finally, an insufficient amount of easily biodegradable carbon (as VFA) in the MBR-Feed on a consistent basis to establish a stable Bio-P removal process. However, there is some optimism with regard to Bio-P removal in an MBR configuration. At an HRT of 6 hours there was evidence of good phosphorus removal during periods of high organic loading. On three occasions total phosphorus levels in the permeate of MBR 3 were below 1 mg/L (0.847, 0.870, and 0.485 mg/L). This is likely due to the shorter SRT than MBR 1 and MBR 2 and the higher organic loading rate to the reactor.

Complete removal of biodegradable carbon was observed, as BOD_5 was consistently lowered below detection limits of 2 mg/L, regardless of influent organic loading. Greater than 93% removal of COD was achieved and is as much dependent on the membrane characteristics as it is on the biological process. A fraction of COD passed through the membrane as supported by the DOC removal results.

Although not representative of the full-scale membrane element, the ZW-10 membrane element performed very well in terms of membrane fouling and low suction pressure requirements. The ZeeWeed® membrane showed no significant increase in suction pressure requirements during the 75-day study. The highly permeable membranes did not require chemical cleaning as a result of membrane fouling and the "relaxation" method of intermittent permeate production proved effective in maintaining stable membrane performance.

6 Recommendations

This project has only touched the surface of MBR research with respect to combined BNR in an MBR configuration. In order to carry on with this work and establish stable and complete removal of nitrogen and phosphorus under typical MBR operating conditions, many recommendations are put forward:

1. A new reactor design should be constructed. The new design should try to achieve an ideal plug-flow configuration to increase t_{10} and prevent nitrate recycle to the anaerobic zone. This includes increasing the number of pre-anoxic cells to provide CFSTRs in series and improve mixing. The reactor should follow a more conventional design with respect to length to width to height dimension ratios. Also the size of the second anoxic zone should be increased to provide a greater hydraulic retention time to remove total oxidized nitrogen in the nitrified mixed liquor recycle.

2. Mixing should be improved in the non-aerated zones without compromising the zero-O_2 environment. At such a small scale, alternatives to mechanical mixing should be explored to eliminate the need for uneconomical specialized motors, bulky motor support structures, shafts and impellers. Miniature submersible pumps may be a practical alternative.

3. The reactor design should provide greater aeration control, especially in the last aerobic zone (cell #6) to prevent dissolved oxygen from being recycled to the anoxic zones and inhibiting denitrification. The aerobic zone could be supplemented with a mechanical mixer or submersible pump to ensure proper mixing with aeration strictly controlled to limit the dissolved oxygen concentration in the last aerobic zone to less than 1.0 mg/L. Online monitoring of oxidation-reduction potential (ORP) at the

entrance and exit of the anoxic and anaerobic zones would provide useful results in determining whether or not optimum conditions exist for denitrification and phosphorus release.

4. Alternative operational control strategies should be investigated. That is, the reactors should be operated at fixed SRTs as opposed to fixed MLSS concentrations. SRTs less than 30 days may improve phosphorus removal due to the physical removal of phosphorus rich biomass from the reactor. Shorter, fixed SRTs will eliminate secondary phosphorus release potential in the aerobic zone. A continuous wasting strategy may also help stabilize the system as opposed to batch wasting.

5. Finally, more frequent zone profiling of the reactors in regards to soluble COD, VFA, TOXN and ortho-phosphate must be undertaken to optimize recycle flow rates and zone sizing. A new reactor design should allow for the adjustment of individual zone volumes to optimize retention times for denitrification and phosphorus release in the unaerated zones.

MBR technology has become a realistic, practical treatment option for municipalities and small communities. The high load capacity, small footprint, and consistent effluent quality without the complexities of clarifier operation make this technology an attractive alternative to CAS systems. Further research at the pilot-scale is necessary in order to assess its feasibility and to determine membrane life expectancy prior to full-scale implementation. This would help determine its true potential in the wastewater treatment industry.

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Appendix A - MBR Control Software

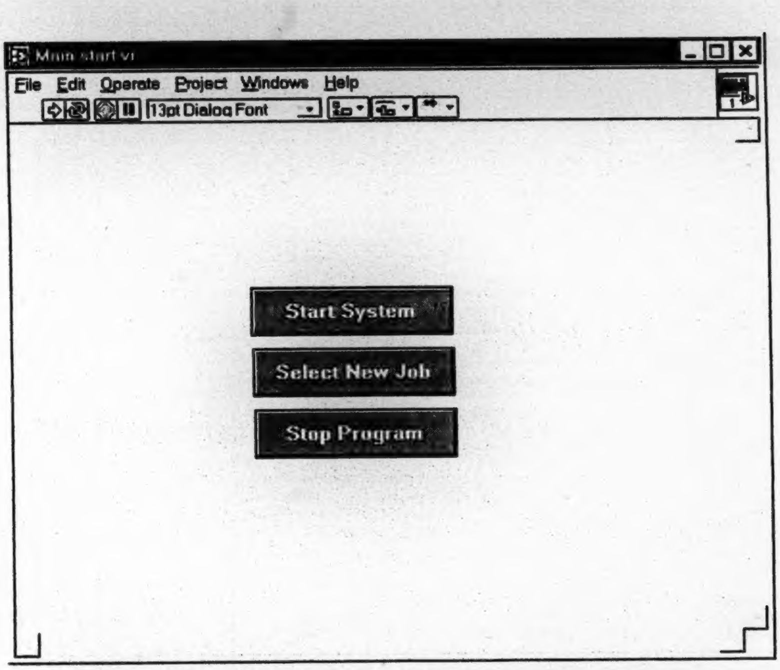
The MBR control software was created by the University of Alberta's Department of Civil and Environmental Engineering using National Instruments Corp. LabVIEW® version 5.1 for Microsoft's® Windows95 operating system.

From the Windows95 desktop, select the "Main Start" icon. The "Main Start" screen appears (Figure 1). To a start a new experimental run and create a new folder, select the "Select New Job" button on the screen. The "Select Job" screen will appear (Figure 2). To create a new job folder, select the "Select Job Name" button. The "New Job Name" screen will appear (Figure 3). Enter the desired job name for the experimental run. A folder will be created with the same name as the job name. Within this folder, four folders will be automatically created. One folder to store data collected from the temperature probe in EXCEL data file format and the other three folders to store data files in EXCEL data format for MBRs 1, 2 and 3. Select the "Exit" button to proceed back to the "Select Job" screen. The newly created job name should be displayed in the "Current Job" box. Past job folders can be by viewed by selecting the "Check Job List". With the current job selected, select the "Exit" button to return to the "Main Start" screen. Select the "Start System" from the "Main Start" screen to proceed to the "Main Control Screen".

The "Main Control Screen" will open in "Manual Mode" (Figure 4). All three MBRs should indicate "Manual" under the "Temperature" display box in the upper left hand corner of the screen. In "Manual Mode", the feed valve and permeate pump are "Off". Temperature will be continually collected in "Manual Mode" with data written to the "Temperature" folder in the current job folder. The "Main Control" screen is divided into four sections, the "Temperature" section, and three "Reactor" sections; one for each MBR. Within the "Temperature" section, up to date temperature statistics collected from the temperature probe, the individual reactor control buttons and a "Stop" button are displayed. The "Stop" button ends all data collection for the current job and returns to the "Main Start" screen. The "Restart" button ends temperature data collection for the current test file, and creates a new one. In this case, "Test#1.dat" will end and "Test#2.dat" will begin if the "Restart" button is selected. As well, a new plot of temperature data will created. For each "Reactor" section, status buttons are displayed for flow conditions and permeate pump operational condition as "On" or "Off". Float status buttons are also displayed along with a "Manual" mode status button. The "Press to Pause Pump" button can be used to manually stop the pump during the "Automatic" mode, but still collect pressure data under the current data file. Under "Manual" mode this button is deactivated because no data is written to files in this mode. The feed valve and permeate pump are controlled via the "Valve On/Off" button and the "Pump On/Off" button in the "Temperature section". These buttons can be used to fill and drain the tank manually without any data collection. Every time the reactors are switched from manual mode to automatic mode a new file called "Test#$_i$.dat" is created within an incremental reference number (i). For example, during start-up, the reactors are in manual mode. After each reactor is switched to automatic mode a new file called "Test#1.dat" is created in each of the reactor folders within the job folder. If the reactors are switched to manual mode, "Test#1.dat" file is closed and "Test#2.dat" file will be created when the reactors are returned to "Automatic" mode. In "Automatic" mode (Figure 5) the MBRs operate as three independent reactors, using the float switches to call for more feed to maintain a desired water level. In "Automatic" mode, the "Manual" status button becomes the "Press to Restart Test" button. This button will stop writing data to the current "Test#$_i$.dat" file and will create a new "Test#$_{i+1}$.dat" data file.

Within the reactor screens, one can set the permeate production and relaxation interval by directly typing in the number of minutes in the "Set Interval Production" boxes. The "Interval Time" screen displays the amount of time passed for each production interval during the run. The "Set MAX TMP" box is used to enter the maximum transmembrane pressure (i.e. suction pressure) allowed (in kilopascals) before the permeate pump is turned off. The pump will immediately enter relaxation mode and remain there until the TMP falls below this set value. The current TMP is displayed on the chart and in the "Current TMP" box. The user can enter the desired frequency of data collection in the "Data every(sec)" box. The TMP data for each reactor is continually charted during automatic mode operation. Typing in the desired maximum TMP directly onto the y-axis can change the y-axis scale on these charts. The same holds true for the continuous temperature chart. The dissolved oxygen data is continually displayed in the "Current DO(mg/L)" box for each reactor.

Table 13 shows an example of a data file collected for one of the three MBRs. The time and date are shown, as well as the temperature, suction pressure, and dissolved oxygen data for probe 1, 2 and 3 (indicated by D1, D2 and D3). Data for all three dissolved oxygen probes are included for each MBR so that one could perform profiling with all three probes in one reactor. In this case, each probe was located in a separate MBR. The float columns indicate the flow status. When all three floats indicate the integer "5", the feed valve is open. When float #3 is close to zero, the feed valve is closed. If float #1 is not equal to "5" then the low-level float in the membrane zone (cell #7) is activated. With this float activated, the permeate pump is deactivated to stop the water level in the tank from dropping below the membrane element header. This safety mechanism prevents the membrane fibres from drying out.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Operate
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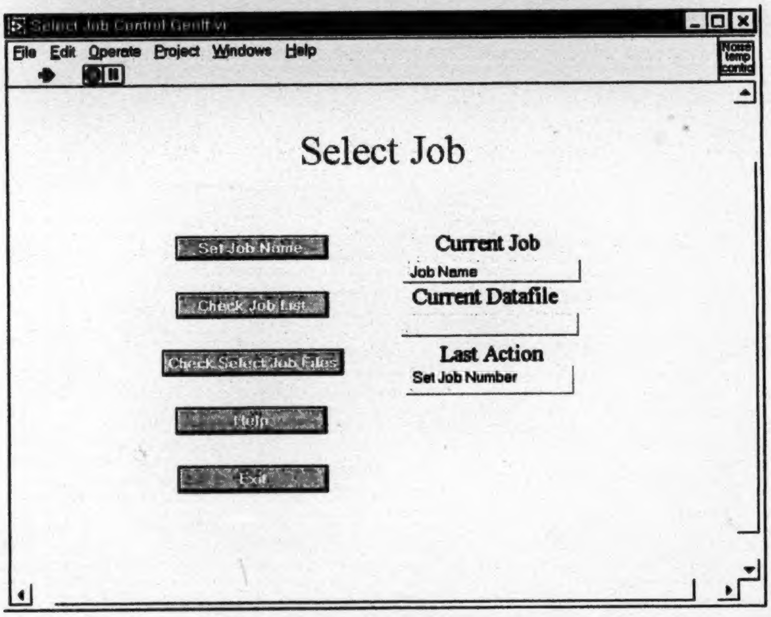
Windows
Help
13pt Dialog Font
Start System
Select New Job
Stop Program

Figure 43. Membrane control software "Main Start" screen.

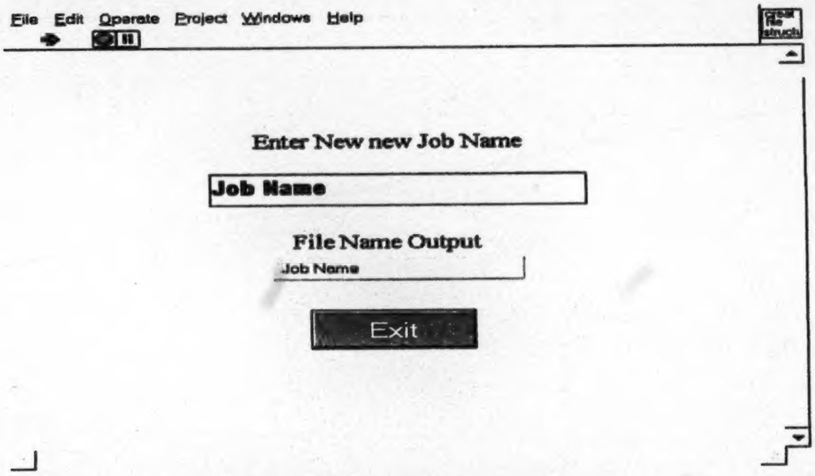

File
Edit
Operate
Project
Windows
Help
Select Job
Set Job Name
Check Job List
Check Select Job Files
Help
Exit

Current Job
Job Name

Current Datafile

Last Action
Set Job Number

Figure 44. "Select Job" screen.


File Edit Operate Project Windows Help
Exit

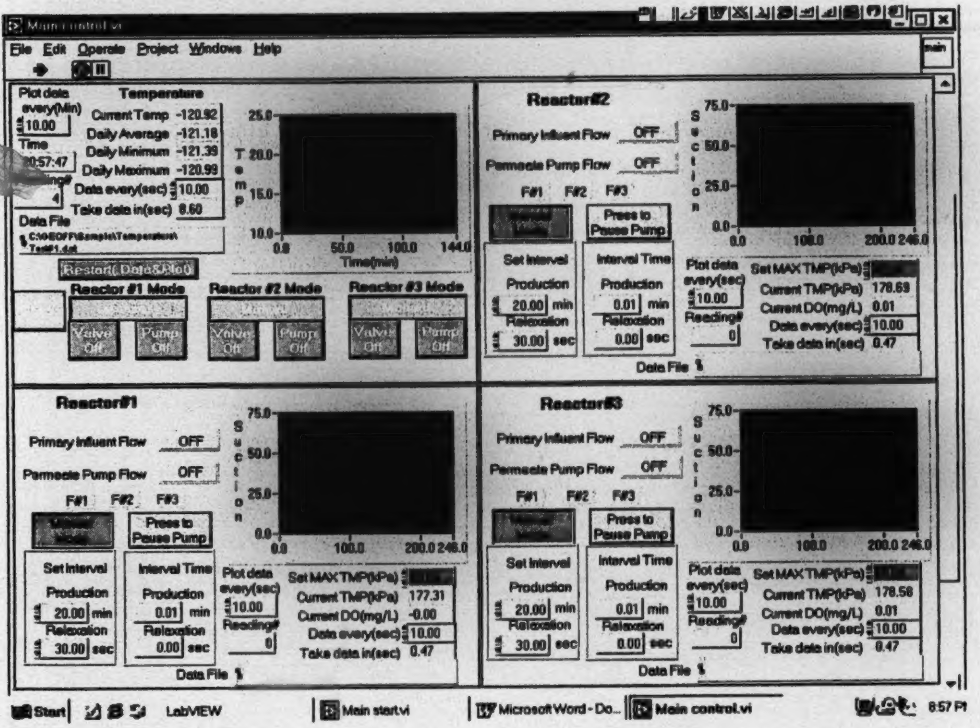
Enter New new Job Name

Job Name

File Name Output

Job Name

Figure 45. "New Job Name" screen.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main
Restart(Data&Plot)
Reactor #1 Mode
Reactor #2 Mode
Reactor #3 Mode
Valve Off
Pump Off
Valve Off
Pump Off
Valve Off
Pump Off
Temp
Time(min)
Suction
F#1 F#2 F#3
Press to Pause Pump
Start
LabVIEW
Main start.vi
Microsoft Word - Do...
Main control.vi
8:57 PM

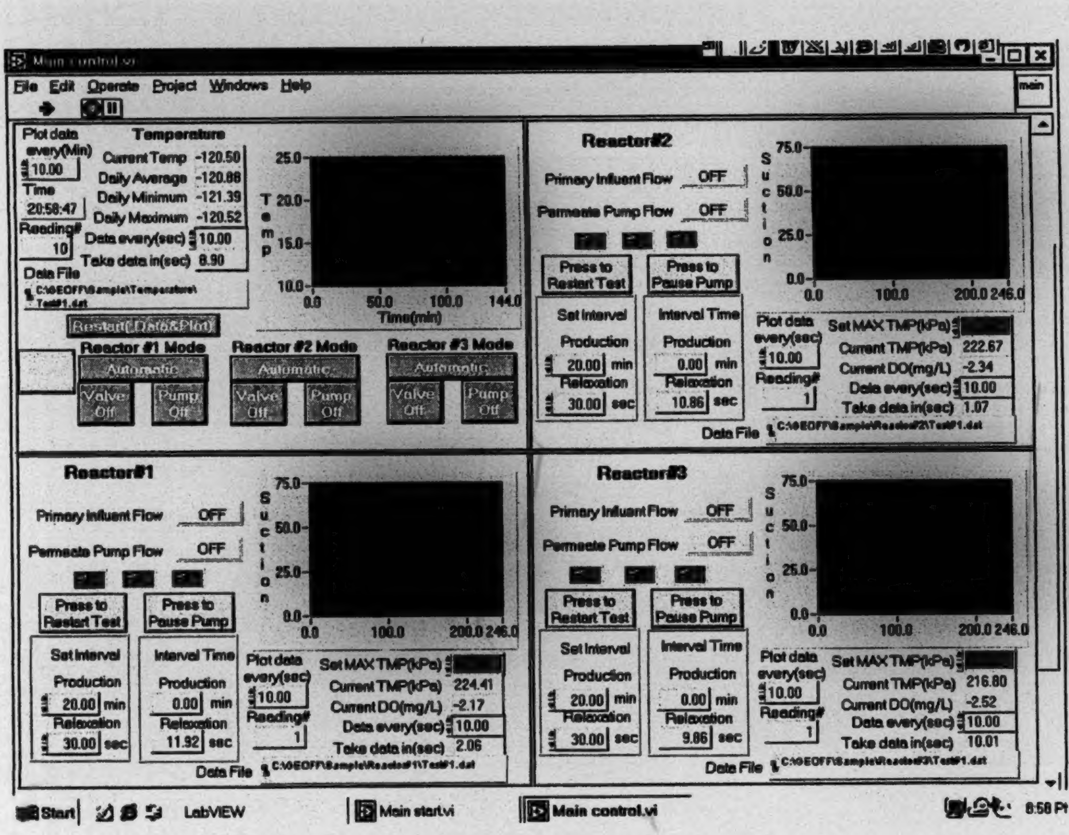
Plot data every(Min) 10.00
Time 20:57:47
Reading# 4
Temperature
Current Temp -120.92
Daily Average -121.18
Daily Minimum -121.39
Daily Maximum -120.99
Data every(sec) 10.00
Take data in(sec) 8.60
Data File C:\OEOFF\Sample\Temperature\Test#1.dat

Reactor#2
Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval Production 20.00 min Relaxation 30.00 sec
Interval Time Production 0.01 min Relaxation 0.00 sec
Plot data every(sec) 10.00
Reading# 0
Set MAX TMP(kPa)
Current TMP(kPa) 178.69
Current DO(mg/L) 0.01
Data every(sec) 10.00
Take data in(sec) 0.47
Data File

Reactor#1
Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval Production 20.00 min Relaxation 30.00 sec
Interval Time Production 0.01 min Relaxation 0.00 sec
Plot data every(sec) 10.00
Reading# 0
Set MAX TMP(kPa)
Current TMP(kPa) 177.31
Current DO(mg/L) -0.00
Data every(sec) 10.00
Take data in(sec) 0.47
Data File

Reactor#3
Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval Production 20.00 min Relaxation 30.00 sec
Interval Time Production 0.01 min Relaxation 0.00 sec
Plot data every(sec) 10.00
Reading# 0
Set MAX TMP(kPa)
Current TMP(kPa) 178.58
Current DO(mg/L) 0.01
Data every(sec) 10.00
Take data in(sec) 0.47
Data File

Figure 46. "Main Control" screen. MBRs 1, 2 and 3 operated in manual mode. Feed valve and permeate pump are "Off".


Main control.vi
File Edit Operate Project Windows Help
main
Temperature
Temp
Time(min)
Restart(Data&Plot)
Reactor #1 Mode
Reactor #2 Mode
Reactor #3 Mode
Automatic
Valve Off
Pump Off
Reactor#2
Reactor#1
Reactor#3
Suction
Press to Restart Test
Press to Pause Pump
Set Interval
Interval Time
Start
LabVIEW
Main start.vi
Main control.vi
8:58 P

Temperature

Plot data every(Min) 10.00
Time 20:58:47
Reading# 10
Current Temp -120.50
Daily Average -120.88
Daily Minimum -121.39
Daily Maximum -120.52
Data every(sec) 10.00
Take data in(sec) 8.90
Data File C:\GEOFF\Sample\Temperature\Test#1.dat

Reactor #1 Mode: Automatic
Reactor #2 Mode: Automatic
Reactor #3 Mode: Automatic

Reactor#2

Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval: Production 20.00 min; Relaxation 30.00 sec
Interval Time: Production 0.00 min; Relaxation 10.86 sec
Plot data every(sec) 10.00
Reading# 1
Current TMP(kPa) 222.67
Current DO(mg/L) -2.34
Data every(sec) 10.00
Take data in(sec) 1.07
Data File C:\GEOFF\Sample\Reactor#2\Test#1.dat

Reactor#1

Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval: Production 20.00 min; Relaxation 30.00 sec
Interval Time: Production 0.00 min; Relaxation 11.92 sec
Plot data every(sec) 10.00
Reading# 1
Current TMP(kPa) 224.41
Current DO(mg/L) -2.17
Data every(sec) 10.00
Take data in(sec) 2.06
Data File C:\GEOFF\Sample\Reactor#1\Test#1.dat

Reactor#3

Primary Influent Flow OFF
Permeate Pump Flow OFF
Set Interval: Production 20.00 min; Relaxation 30.00 sec
Interval Time: Production 0.00 min; Relaxation 9.86 sec
Plot data every(sec) 10.00
Reading# 1
Current TMP(kPa) 216.80
Current DO(mg/L) -2.52
Data every(sec) 10.00
Take data in(sec) 10.01
Data File C:\GEOFF\Sample\Reactor#3\Test#1.dat

Figure 47. "Main Control" screen with MBR1, 2 and 3 operated in "Automatic" mode.

Table 13. Example of online data collected for MBR.

Time	Date	Run#	Temp	Pressure	D1	D1	D1	Float#1	Float#2	Float#3
1:42:39 PM	8/30/01	100	-55.82	5.54	4.25	4.29	4.56	5	5	5
1:43:39 PM	8/30/01	101	-56.44	5.6	3.56	4.3	4.05	5	5	0.002
1:44:39 PM	8/30/01	102	-56.44	5.6	4.25	4.29	4.17	5	5	0.002
1:45:39 PM	8/30/01	103	-56.44	5.58	4.18	4.3	4.13	5	5	0.002
1:46:39 PM	8/30/01	104	-55.83	5.64	4.19	4.27	3.93	5	5	5
1:47:39 PM	8/30/01	105	-56.43	5.6	4.27	4.28	4.16	5	5	0.002
1:48:39 PM	8/30/01	106	-56.43	5.62	3.78	4.28	4.29	5	5	0.002
1:49:40 PM	8/30/01	107	-56.43	5.61	4.31	4.28	4.13	5	5	0.002
1:50:40 PM	8/30/01	108	-56.43	5.59	4.06	4.27	4.09	5	5	0.002
1:51:39 PM	8/30/01	109	-56.43	5.59	3.75	4.3	4.11	5	5	0.002
1:52:39 PM	8/30/01	110	-56.43	5.6	4.22	4.28	3.99	5	5	0.002
1:53:39 PM	8/30/01	111	-56.43	5.56	4.2	4.29	3.99	5	5	0.002
1:54:39 PM	8/30/01	112	-55.83	5.6	4.06	4.27	4.08	5	5	5

Appendix B - Pressure Transducer Calibration

Water Temperature	23	^{o}C
Density @ 23^{o}C	0.99756	g/mL
Density @ 20^{o}C	0.99823	g/mL

Height (inches H_20)	Voltage (Volts)	Height (corrected to 20^{o}C)	Pressure (kPa)
85	2.59	84.94	21.2
65.5	2.52	65.46	16.3
48.75	2.47	48.72	12.1
33	2.42	32.98	8.2
17.25	2.37	17.24	4.3

Figure 48. Calibration results for pressure transducer #1.

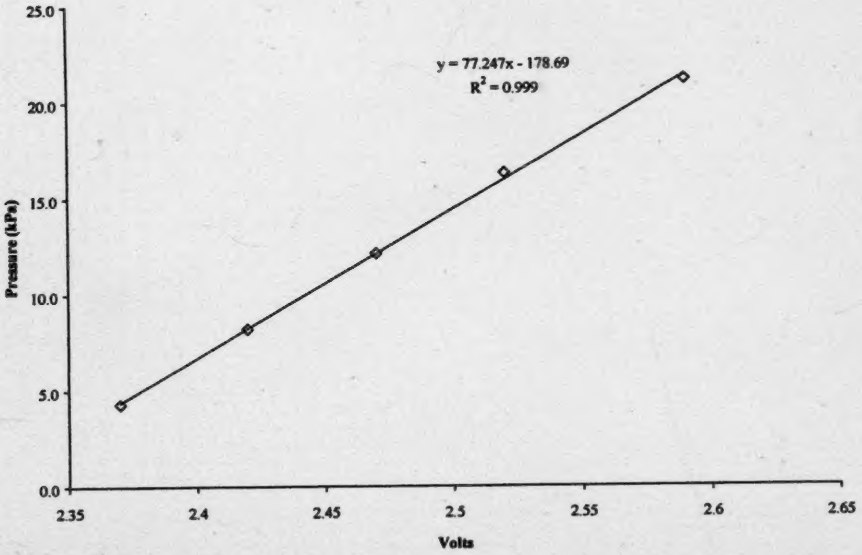


Figure 49. Plot of pressure (kPa) versus volts for pressure transducer #1. Linear equation was generated using Microsoft® EXCEL's linear regression tool.

Water Temperature	20.5	°C
Density @ 21°C	0.99802	g/mL
Density @ 20°C	0.99823	g/mL

Height (inches H_2O)	Voltage (Volts)	Height (corrected to 20°C)	Pressure (kPa)
84.25	2.58	84.23	21.0
65.5	2.52	65.49	16.3
49.5	2.47	49.49	12.3
33.25	2.42	33.24	8.3
17	2.36	17.00	4.2

Figure 50. Calibration results for pressure transducer #2.

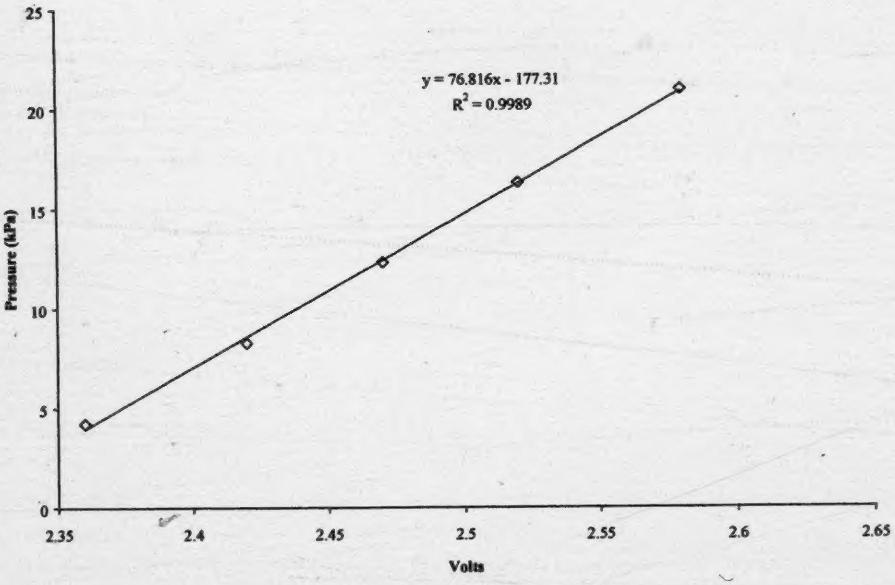


Figure 51. Plot of pressure (kPa) versus volts for pressure transducer #2. Linear equation was generated using Microsoft® EXCEL's linear regression tool.

Water Temperature	14.9	°C	
Density @ 15°C	0.99913	g/mL	
Density @ 20°C	0.99823	g/mL	
Height (inches H$_2$0)	Voltage (Volts)	Height (corrected to 20°C)	Pressure (kPa)
84	2.58	84.08	20.9
67	2.53	67.06	16.7
49.5	2.47	49.54	12.3
33.5	2.42	33.53	8.3
17	2.365	17.02	4.2

Figure 52. Calibration results for pressure transducer #3.

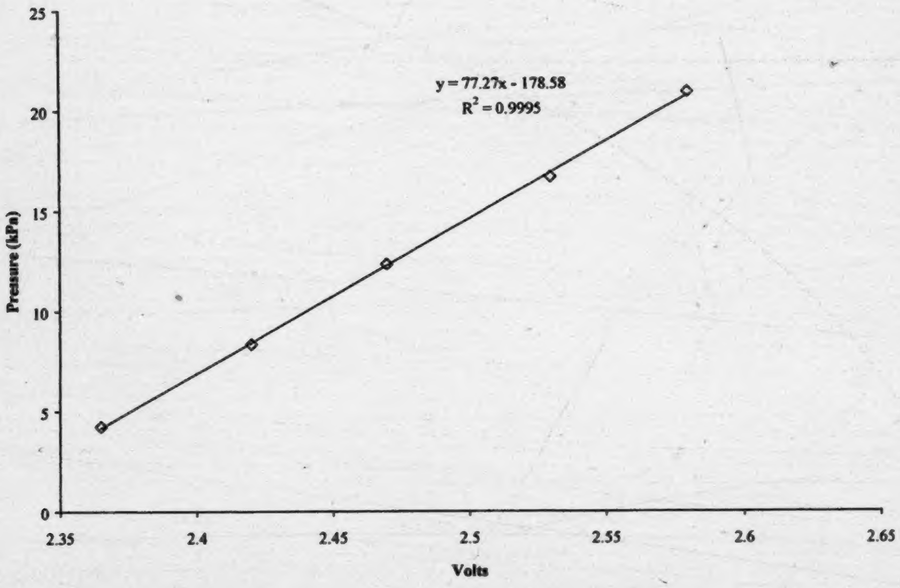


Figure 53. Plot of pressure (kPa) versus volts for pressure transducer #3. Linear equation was generated using Microsoft® EXCEL's linear regression tool.